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Economic Evaluation of a Sodium/Limestone Double-Alkali FGD (Flue Gas Desulfurization) Process

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by

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16. ABSTRACT The report gives results of a comparison of results from a recent forced-oxidation limestone flue gas desulfurization (FGD) process evaluation and those from a conceptual design and economic evaluation of a sodium/limestone double-alkali FGD process, based on recent EPA-sponsored pilot-plant and prototype test work. For a 500-MW new power unit burning 3.5% sulfur coal and meeting the 1979 new source performance standards, the estimated capital investments in 1982 costs are $95 million ($190/kW) for the sodium/limestone double-alkali process and $103 million ($206/kW) for the forced-oxidation limestone process. Estimated first-year annual revenue requirements in 1984 costs for these processes are $26 million and $29 million (9.3 and 10.6 mills/kWh), respectively. Although the sodium/limestone double-alkali process appears to be about 8% lower in capital investment, given the accuracy associated with studies of this type (+ or - 10%), it is uncertain if the sodium/limestone double-alkali process has a lower capital investment. In terms of first-year and levelized annual revenue requirements, the sodium/limestone double-alkali process shows a 12% and 14% lower cost, respectively, than the forced-oxidation limestone process and (thus) is marginally less expensive. However, economics for the double-alkali process are more uncertain than those for the other.

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a. DESCRIPTORS		b. IDENTIFIERS/OPEN ENDED TERMS	c. COSATI Field/Group
Pollution	Alkalies	Pollution Control	13B
Flue Gases	Coal	Stationary Sources	21B 21D
Desulfurization	Combustion	Sodium/Limestone	07A, 07D
Scrubbers		Double-Alkali Process	13I
Sodium			07B
Calcium Carbonates			

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ABSTRACT

A conceptual design and economic evaluation of a sodium/limestone double-alkali flue gas desulfurization (FGD) process was prepared based on recent EPA-sponsored pilot-plant and prototype test work. The results of this evaluation were compared with the results from a recent forced-oxidation limestone process evaluation. For a 500-MW new power unit burning 3.5% sulfur coal and meeting the 1979 new source performance standards (NSPS), the estimated capital investments in 1982 costs are 95 M$ (190 $/kW) for the sodium/limestone double-alkali process and 103 M$ (206 $/kW) for the forced-oxidation limestone process. Estimated first-year annual revenue requirements in 1984 costs for these processes are 26 M$ and 29 M$ (9.3 and 10.6 mills/kWh), respectively. Although the sodium/limestone double-alkali process appears to be about 8% lower in capital investment, given the accuracy associated with studies of this type (±10%), it is uncertain whether the sodium/limestone double-alkali process has a lower capital investment. In terms of first-year and levelized annual revenue requirements, the sodium/limestone double-alkali process shows a 12% and 14% lower cost than the forced-oxidation limestone process and thus is marginally less expensive. However, some of the design assumptions used to generate the estimated costs for the sodium/limestone double-alkali process are based on short-term pilot-plant tests and are unconfirmed. Therefore, the economics are more uncertain for the double-alkali process than for the forced-oxidation limestone process.

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ABBREVIATIONS AND CONVERSION FACTORS

ABBREVIATIONS

aft3/min	actual cubic feet per minute	lb	pound
Btu	British thermal unit	L/G	liquid-to-gas ratio in gallons per thousand actual cubic feet of gas at outlet conditions
oF	degrees Fahrenheit		
dia	diameter		
FGD	flue gas desulfurization	M	million
ft	feet	mi	mile
ft2	square feet	mo	month
ft3	cubic feet	MW	megawatt
gal	gallon	ppm	parts per million
gpm	gallons per minute	psig	pounds per square inch (gauge)
gr	grain		
hp	horsepower	rpm	revolutions per minute
hr	hour	sec	second
in.	inch	sft3/min	standard cubic feet per minute (60oF)
k	thousand		
kW	kilowatt	SS	stainless steel
kWh	kilowatthour	yr	year

CONVERSION FACTORS

EPA policy is to express all measurements in Agency documents in metric units. Values in this report are given in British units for the convenience of engineers and other scientists accustomed to using the British systems. The following conversion factors may be used to provide metric equivalents.

To convert British		Multiply by	To obtain Metric	
ac	acre	0.405	hectare	ha
Btu	British thermal unit	0.252	kilocalories	kcal
^{o}F	degrees Fahrenheit minus 32	0.5556	degrees Celsius	^{o}C
ft	feet	30.48	centimeters	cm
ft^2	square feet	0.0929	square meters	m^2
ft^3	cubic feet	0.02832	cubic meters	m^3
ft/min	feet per minute	0.508	centimeters per second	cm/s
ft^3/min	cubic feet per minute	0.000472	cubic meters per second	m^3/s
gal	gallons (U.S.)	3.785	liters	L
gpm	gallons per minute	0.06308	liters per second	L/s
gr	grains	0.0648	grams	g
gr/ft^3	grains per cubic foot	2.288	grams per cubic meter	g/m^3
hp	horsepower	0.746	kilowatts	kW
in.	inches	2.54	centimeters	cm
lb	pounds	0.4536	kilograms	kg
lb/ft^3	pounds per cubic foot	16.02	kilograms per cubic meter	kg/m^3
lb/hr	pounds per hour	0.126	grams per second	g/s
psi	pounds per square inch	6895	pascals (newton per square meter)	Pa (N/m^2)
mi	miles	1609	meters	m
rpm	revolutions per minute	0.1047	radians per second	rad/s
sft^3/min	standard cubic feet per minute ($60^{o}F$)	1.6077	normal cubic meters per hour ($0^{o}C$)	m^3/hr ($0^{o}C$)
ton	tons (short)	0.9072	metric tons	tonne
ton/hr	tons per hour	0.252	kilograms per second	kg/s

ECONOMIC EVALUATION OF A SODIUM/LIMESTONE DOUBLE-ALKALI FGD PROCESS

EXECUTIVE SUMMARY

For the past 15 years, wet limestone- and lime-scrubbing processes have in almost all cases held a distinct cost advantage over other flue gas desulfurization (FGD) processes. However, limestone and lime processes have several inherent disadvantages, i.e., erosion, corrosion, and a high liquid recirculation rate. They were also plagued during the course of their early development by operating problems such as scaling and plugging. These various shortcomings have added impetus to the continuing development of alternative FGD processes.

One such alternative is the use of double-alkali (dual alkali) processes in which the absorption function is separated from the precipitation function. A highly reactive absorbent solution is used to absorb SO_2, which is precipitated as $CaSO_3 \cdot 1/2H_2O$ and $CaSO_4 \cdot 2H_2O$ outside the absorber. The waste produced is thus identical to that produced by limestone and lime processes but the difficulties of scrubbing with a calcium-based slurry are avoided.

During the course of the development of double-alkali processes in the United States, a solution of sodium sulfite has been used as the absorbent and lime has been used as the precipitant. The use of lime, which is due to its high reactivity, has come to be regarded as a disadvantage because of its cost. One phase of a continuing Environmental Protection Agency (EPA) double-alkali development program has been devoted to the development of a double-alkali process using limestone instead of lime. That process was tested recently on a pilot scale at Gulf Power Company's Scholz Power Station. The results of that test program are subject to question because the pilot plant did not achieve a long-term, closed-loop operation.

This study is an economic evaluation of the sodium/limestone double-alkali process and an economic comparison of the sodium/limestone double-alkali process with a conventional forced-oxidation limestone-scrubbing process.

DESIGN AND ECONOMIC PREMISES

The major design premises for this study are listed in Table S-1. The base case power plant is a new, 500-MW coal-fired power unit located in the

north-central region of the United States. The fuel is a bituminous coal with a heating value of 11,700 Btu/lb and containing 3.5% sulfur (dry basis), 15.14% ash, and 4.0% moisture. The boiler heat rate is 9,500 Btu/kWh.

TABLE S-1. MAJOR DESIGN PREMISES

Item	Premise
Power plant	North-central region, 500-MW coal-fired boiler, 9,500 Btu/kWh heat rate
Operating schedule	165,000 hr, 30-yr life, 5,500-hr first-year operation
Fuel	Eastern bituminous coal, 11,700 Btu/lb, 3.5% sulfur, 15.14% ash, 4.0% moisture
Base year	Capital investment: mid-1982 Revenue requirements: 1984
FGD waste disposal	Sodium/limestone double-alkali fixation and landfill
	Forced-oxidation limestone scrubbing: landfill
SO_2 removal efficiency	89%
Particulate removal efficiency	99.8% (0.03 lb of particulates/MBtu heat input)
SO_2 absorber redundancy	25% (4 operating trains, 1 spare)

The FGD unit includes all the equipment necessary to meet the 1979 new source performance standards (NSPS) for SO_2 removal. The overall design for the sodium/limestone double-alkali process is based on pilot-plant studies at the Scholz Power Station and bench-scale work at the Industrial Environmental Research Laboratory (IERL), Research Triangle Park (RTP), North Carolina. The design of the forced-oxidation limestone process is based on industry data and previous evaluations by TVA.

The project is assumed to begin in mid-1980 with a 3-year construction period ending in mid-1983. The midpoint for the capital investment costs is mid-1982. The annual revenue requirements are based on 1984 costs.

Capital investment consists of direct investment, indirect investment, and other capital charges. The direct investment is based on equipment costs and installation costs (such as piping, electrical, instrumentation, etc.). Indirect investment (engineering design and supervision, construction expense, etc.) is estimated based on the direct investment. Other capital costs (allowance for startup and modification, interest during construction, etc.) are estimated from the total direct and indirect investment. These preliminary capital investment estimates are considered to have a -15% to +30% range of accuracy and to be comparable within 10%.

Two types of annual revenue requirements are projected--first year and levelized. Both are based on 5,500 hours of operation per year at full load (about a 63% capacity factor) and both use a levelized capital charge. However, levelized annual revenue requirements differ from first-year annual revenue requirements in that they take into consideration the time value of money over the life of the FGD unit and are calculated using a 10% discount factor, 6% inflation factor, and a 30-year economic life.

PROCESS DESCRIPTIONS

The sodium/limestone double-alkali process (Figure S-1) is similar in design and operation to utility and industrial sodium/lime double-alkali processes except for the additional reaction tanks and the limestone-grinding equipment. The forced-oxidation limestone process (Figure S-2) is a generic design based on current technology, including the use of forced oxidation to produce gypsum. Both systems have five absorber trains--four operating and one spare--supplied with flue gas from a plenum downstream from the electrostatic precipitator (ESP) and the boiler induced-draft (ID) fan. In both cases, ducts for emergency bypass of 50% of the flue gas are provided. Process design conditions for both systems are summarized in Table S-2.

Each train of the sodium/limestone double-alkali system consists of a sieve tray absorber, a steam reheater, and a booster fan. The absorbent liquid is a sodium sulfite solution containing 1.7 mol percent sodium and 0.7 mol percent active alkali (SO_3^{2-} and HSO_3^-) when regenerated. Bleedstreams from the absorbers are combined and treated with ground limestone in two parallel trains of four reaction tanks operating in an overflow mode. The resulting slurry is pumped to a thickener-filter train where a 55% solids filter cake is produced. The filter cake is blended with dry fly ash and lime and trucked one mile to a landfill.

Each train of the forced-oxidation limestone process consists of a spray tower absorber, a steam reheater, and a booster fan. The absorbent liquid is an 8% solids slurry containing ground limestone. The scrubbing slurry drains from the absorber into a tank where air is sparged into it to oxidize the

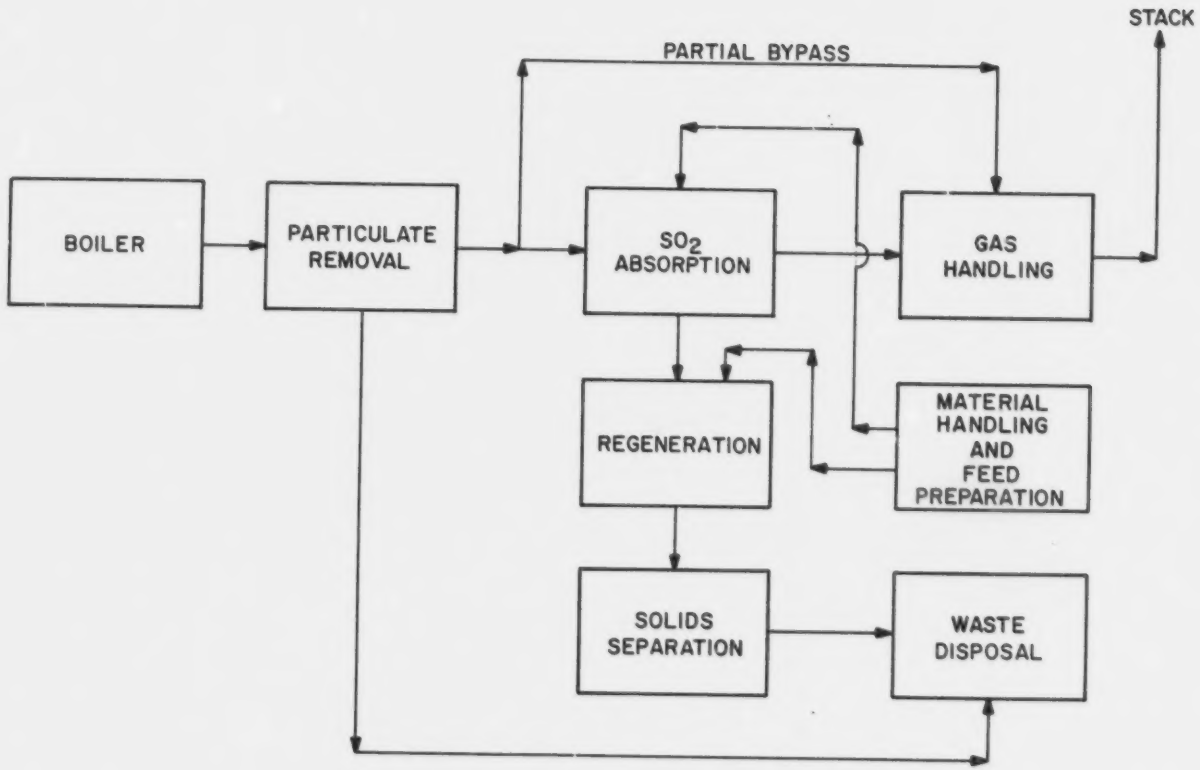

Figure S-1. Sodium/limestone double-alkali process flow diagram.

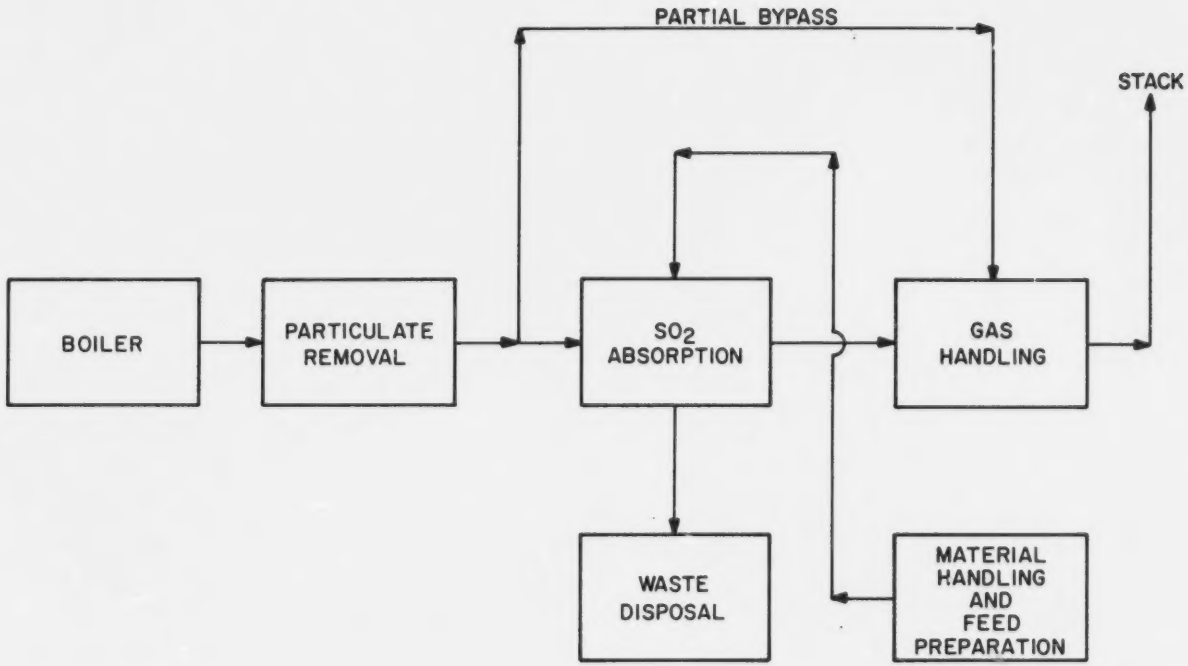

Figure S-2. Forced-oxidation limestone process flow diagram.

sulfite to sulfate before it overflows to a hold tank where the makeup limestone slurry is added. A bleedstream from the oxidation tank is pumped to a thickener-filter train where an 85% solids filter cake is produced. The filter cake is then trucked one mile to a landfill.

TABLE S-2. PROCESS DESIGN CONDITIONS

	Process	
	Sodium/limestone double alkali	Limestone scrubbing
Absorber type	Sieve tray tower	Spray tower
Superficial gas velocity, ft/sec	9	10
L/G, gal/kaft^3		
Presaturator/underspray	2/1	4/0
Absorber	5	106
Stoichiometry, mol Ca/mol (SO_2 + 2HCl) absorbed	1.0	1.4
Soda ash feed rate, mol Na^+/mol SO_2 absorbed	0.07	-
Sulfite oxidation, %	10	95
Thickener feed solids, %	1.4	8
Thickener underflow solids, %	25	40
Filter cake solids, %	55	85

RESULTS

The capital investments and annual revenue requirements of the sodium/limestone double-alkali process and the forced-oxidation limestone process are shown in Table S-3. Fly ash disposal costs are not included in the forced-oxidation limestone process costs. Since fly ash disposal is an integral part of the fixation portion of the sodium/limestone double-alkali process, a credit for fly ash disposal is applied to the costs of the sodium/limestone double-alkali process for comparability with the forced-oxidation limestone process costs. The capital investment and annual revenue requirements of the sodium/limestone double-alkali process are both lower than those of the

forced-oxidation limestone process--by 8% and 12%, respectively. Since the comparative accuracy of these cost estimates is 10%, the capital investment of the double-alkali process can be regarded as marginally less expensive.

TABLE S-3. COMPARISON OF CAPITAL INVESTMENTS AND ANNUAL REVENUE REQUIREMENTS

	Sodium/limestone double-alkali process	Forced-oxidation limestone process
Capital investment (1982 $)		
M$	95.2	103.1
$/kW	190.3	206.1
First-year revenue requirements (1984 $)		
M$	25.5	29.1
Mills/kWh	9.3	10.6
Levelized annual revenue requirements (1984 $)		
M$	35.7	41.5
Mills/kWh	13.0	15.1

Capital Investment

The direct capital investments (the cost of material and labor to install the system) are compared by process area in Table S-4. It is apparent that the key to the economic advantage of the sodium/limestone double-alkali process is the less expensive SO_2 absorption area that results from the reduced pumping requirements made possible by the low liquid-to-gas (L/G) ratio (the absorber sizes are similar because the size is determined largely by the flue gas volume). The SO_2 absorption area costs for the double-alkali process are one-half those for the limestone process and constitute 24% of the total direct investment, while for the limestone process, they constitute 36% of the total direct investment. Cost differences in other processing areas are a much lower portion of the total. Regeneration area costs (3%) and fixation area costs (2%) for the sodium/limestone double-alkali process are minor, as are oxidation area costs (5%) for the forced-oxidation limestone process. The solids separation area (dewatering) costs for the double-alkali process are 50% higher but these, too, are a relatively small part of the total direct investment--12% for the sodium/limestone double-alkali process and 7% for the forced-oxidation limestone process.

TABLE S-4. COMPARISON OF DIRECT INVESTMENT BY PROCESSING AREA

Processing area	Direct investment, 1982 k$ Sodium/limestone double-alkali process	Forced-oxidation limestone process
Materials handling	2,426	2,528
Feed preparation	4,506	4,715
Gas handling	10,800	11,281
SO_2 absorption	11,348	20,288
Reheat	3,630	3,634
Regeneration	1,506	-
Oxidation	-	2,670
Solids separation	5,493	3,679
Fixation	906	-
Services, utilities, and misc.	2,437	2,928
Landfill construction	5,247	3,781
Landfill equipment	1,454	1,123
Landfill credit (fly ash disposal)	(2,312)	-
Total	47,441	56,627

Annual Revenue Requirements

A comparison of the major components of the first-year annual revenue requirements is shown in Table S-5. Among the direct costs, the major costs for the sodium/limestone double-alkali process, in order of importance, are maintenance, raw materials, labor, steam, and electricity. For the forced-oxidation limestone process, they are maintenance, electricity, steam, labor, and raw materials. The major differences in direct costs between the processes are in maintenance, electricity, and raw material costs. Maintenance costs for the sodium/limestone double-alkali process are 37% lower than for the forced-oxidation limestone process as a result of the forced-oxidation limestone process having high maintenance costs associated with circulating large volumes of slurry. This also results in electricity costs for the sodium/limestone double-alkali process being about half those of the forced-oxidation limestone process. Raw material costs are over twice as high for the sodium/limestone double-alkali process, however, and are divided about equally between costs for limestone, soda ash, and lime for fixation.

Overall, the sodium/limestone double-alkali process direct costs are about 13% lower than those of the forced-oxidation limestone process because of lower maintenance and electricity costs, both a result of using a highly

reactive scrubbing solution instead of a limestone slurry. With the lower overheads, which are based on direct costs, and lower capital charges, the annual revenue requirements of the sodium/limestone double-alkali process are 12% lower than those of the forced-oxidation limestone process.

TABLE S-5. COMPARISON OF ANNUAL REVENUE REQUIREMENT COMPONENTS

	Annual revenue requirements, 1984 k$	
	Sodium/limestone double-alkali process	Forced-oxidation limestone process
Raw materials		
Limestone	915	1,214
Soda ash	990	-
Lime	794	-
Total labor	1,357	1,269
Utilities		
Steam	1,321	1,356
Electricity	1,096	2,133
Other	234	192
Maintenance	2,715	4,285
Landfill credit	(312)	-
Analysis	105	105
Direct costs	9,215	10,554
Overheads	2,319	3,395
Operating and maintenance costs	11,534	13,949
Levelized capital charges	13,987	15,151
Total first-year annual revenue requirements	25,521	29,100

Energy Consumption

A comparison of the total energy consumption for each process is shown in Table S-6. Steam, diesel fuel, and electricity are shown in Btu equivalents.

TABLE S-6. COMPARISON OF PROCESS ENERGY CONSUMPTION

Process	Energy input, 10^9 Btu equivalent			% of plant capacity (gross)
	Electricity[a]	Steam[b]	Diesel fuel[c]	
Sodium/limestone double alkali	281.4	397.3	11.6[d]	2.6
Limestone scrubbing	547.7	407.7	14.9	3.7

a. 9,500 Btu/kWh.
b. 751.9 Btu/lb (based on recycle of condensate to power plant).
c. 144,000 Btu/gal.
d. Includes credit for fly ash disposal.

The energy consumption in the sodium/limestone double-alkali process is only 71% of that for the forced-oxidation limestone process. Almost all of the difference is due to the significantly lower electrical consumption of the sodium/limestone double-alkali process.

CONCLUSIONS AND RECOMMENDATIONS

The sodium/limestone double-alkali process is 8% lower in capital investment, 12% lower in first-year annual revenue requirements, and 14% lower in levelized annual revenue requirements than a comparable forced-oxidation limestone process. Although these economics may be only marginally attractive relative to limestone scrubbing (given the accuracy associated with this preliminary-grade economic evaluation), the sodium/limestone double-alkali process offers the potential for high SO_2 removal efficiencies at low absorber L/G ratios and for fewer maintenance problems with scaling, plugging, and erosion because calcium compounds are primarily confined to the regeneration area.

The sodium/limestone double-alkali process is less energy intensive than a forced-oxidation limestone process--primarily because of the lower electrical consumption of recirculating pumps in the scrubbing area.

More development work on the sodium/limestone double-alkali process needs to be undertaken to demonstrate that both the design assumptions used in this study are achievable and the problems associated with a previous pilot-plant work at Scholz can be overcome with only minor (and relatively inexpensive)

modifications. Both of these areas of uncertainty could be minimized with additional pilot-plant or prototype (3- to 30-MW equivalent) work which includes a long-term demonstration run. As a result of the unconfirmed design assumptions mentioned above, the estimated economics are considered potentially less accurate for the double-alkali process than for the forced-oxidation limestone process.

ECONOMIC EVALUATION OF A SODIUM/LIMESTONE DOUBLE-ALKALI FGD PROCESS

INTRODUCTION

This report presents the results of a conceptual design and economic evaluation of a sodium/limestone double-alkali flue gas desulfurization (FGD) process. The process uses a sodium-based scrubbing solution. The spent absorbent is withdrawn from the absorber and treated with limestone to regenerate the scrubbing solution and produces a mixture of calcium-sulfur salts, which are fixed with fly ash and lime and disposed of in a landfill.

This evaluation is part of a continuing series of economic evaluations of FGD processes performed by the Tennessee Valley Authority (TVA) for the U.S. Environmental Protection Agency (EPA). These evaluations were begun in 1967 and have focused both on current and emerging FGD technology. During this period, both regulatory standards governing sulfur oxides (SO_x) emissions and the FGD technology to meet the regulations have evolved. Similarly, TVA's economic evaluation procedures have evolved and broadened to adapt to the demands of legislative, technical, and economic changes.

Double-alkali (also called dual alkali) processes are a large and heterogeneous class of FGD processes in which the absorption step is separate from the precipitation step to some degree. In some double-alkali processes, the objective is to obtain a low pH absorbent liquid for in-loop forced oxidation. In most cases, however, the objective is to obtain a more desirable absorbent liquid than a lime or limestone slurry, while still producing a calcium salt product. The most numerous and clearly defined double-alkali processes are those in which a solution is used as the scrubbing liquid and a bleedstream is treated with lime or limestone to precipitate calcium sulfur salts and regenerate the absorbent for reuse. The product can be either a high-sulfite sludge or, if forced oxidation is used, gypsum. Several types of scrubbing solutions are used, including those based on ammonia, acetate, and chloride salts, but most use a sodium-based solution prepared from soda ash (sodium carbonate) in which the sulfite ion (SO_3^{2-}) is the active alkaline species. This class of double-alkali processes is the subject of this study.

The use of a highly reactive sodium-based solution avoids the corrosion and plugging problems of calcium-based slurries and reduces the gypsum scaling problems associated with calcium-based slurries. It also permits the use of

lower liquid-to-gas (L/G) ratios than those required for limestone processes. Against these advantages, however, must be weighed the added complexity of the process. Also, there has been a tendency, particularly in the United States, to use more reactive lime (sodium/lime processes) rather than limestone (sodium/limestone processes) as the precipitant, which substantially increases operating costs. A major objective of this study is the evaluation of the economics of the use of limestone rather than lime.

Double-alkali processes have been most widely used in Japan where 51 systems (representing 16% of the total flue gas volume scrubbed) were in use in 1980 (1). All of these processes produced gypsum. In fact, this wide use of double-alkali technology and the diverse design concepts represented are a result of the Japanese emphasis on byproduct production. These processes have been well described in the literature, particularly by Ando (2) in a series of summations of Japanese FGD technology at EPA FGD symposiums. Most FGD applications in Germany use the Saarberg-Holter double-alkali process (3).

Sodium/lime double-alkali processes have been popular in industrial applications in the United States but they have found limited utility application. Three double-alkali systems, all started up in 1979 and all sodium/lime processes, are in operation on utility boilers and two others are contracted. These are discussed in the background section.

EPA has provided a major impetus to the development of double-alkali processes in the United States through a planned program of laboratory and pilot-plant studies and support of prototype and full-scale evaluations. This work has been widely reported and summarized in the literature, much of it in the proceedings of EPA FGD symposiums (4). The program, which began in 1971 to provide an alternative to the problem-ridden lime and limestone FGD processes of the time, is discussed in the background section.

Initially, the EPA program focused on sodium/lime processes. By 1979 (when the first full-scale utility sodium/lime system sponsored by EPA was started up at the Louisville Gas and Electric Company's Cane Run Station), attention had turned to other areas of double-alkali technology. One of these areas was the use of limestone instead of lime as the precipitant, a major result of which would be to make the double-alkali process directly comparable with the widely used conventional limestone process. In a previous TVA economic comparison of a sodium/lime double-alkali process and a conventional limestone process (5), the cost of lime was found to be a major operating cost. In spite of this, the double-alkali process was only marginally more expensive to operate than the conventional limestone process. The use of limestone rather than lime could thus have a significant effect on the economic relationships of the two processes. In addition, the study was based on regulatory and economic conditions and FGD technology that prevailed in the late 1970s [1971 new source performance standards (NSPS), late 1970 cost projections, and pond disposal, for example] and does not fully represent current practices.

This evaluation compares the economics of a sodium/limestone double-alkali process with the economics of a conventional forced-oxidation limestone

process under conditions projected to the mid-1980s. The double-alkali process is based on data developed by EPA, as discussed in the background section. In addition to illustrating the economic effects of using limestone instead of lime, it provides an updated evaluation based on current regulatory conditions, FGD technology, and economics. The economic evaluation consists of the capital investment and annual revenue requirements for a 500-MW, 3.5% sulfur eastern bituminous coal application. The process economics for a conventional forced-oxidation limestone process are included for comparison.

BACKGROUND

In spite of the widespread application of wet-limestone scrubbing for utility FGD applications, there are several inherent disadvantages in limestone scrubbing (e.g., scaling, corrosion, plugging, and high L/G ratios) which have provided impetus for continued research and development of economical alternative FGD processes. One class of alternative FGD processes, the sodium/lime double-alkali processes, has received considerable attention during the past decade. The use of a sodium-based scrubbing solution results in higher SO_2 removal efficiencies at lower absorber L/G ratios and minimizes scaling, plugging, and corrosion in the absorption area by confining the calcium-based slurry to the regeneration and disposal areas. Lime has been favored for regeneration of the sodium-scrubbing solution because of its high reactivity. The feasibility of using limestone was questioned because of its low reactivity, which could necessitate more complex process equipment and result in environmentally unsound waste containing sodium.

PROCESS CHEMISTRY

The chemistry of sodium-based double-alkali processes is appreciably less complicated than the chemistry of limestone and lime slurry processes at equivalent levels of sophistication, partly because the solid-liquid reactions are confined to the regeneration process outside the absorber and partly because a single anion provides almost all of the effective alkali. The chemistry has been widely discussed in the literature (6,7,8,9)--much of it in terms of sodium/lime processes, although the chemistry of sodium/limestone processes is quite similar (10). The essential chemical reactions for either the sodium/lime or the sodium/limestone processes can, in fact, be defined by two equations, one for the absorption reaction and one for the regeneration reaction.

Definitions

Several terms used in discussions of sodium-based double-alkali chemistry can be baffling without an understanding of their meanings (9). "Active alkali" is the concentration of anionic species that are basic to dissolved SO_2, or which can be made so in the regeneration process (these are normally SO_3^{2-}, HSO_3^-, OH^-, CO_3^{2-}, and HCO_3^-). The "molar capacity" of the scrubbing solution is the number of mols of SO_2 that can be converted to HSO_3^- by an average mol of active alkali. For example, the molar capacity of CO_3^{2-} is 2, that of SO_3^{2-} is 1, and that of HSO_3^- is 0. The molar capacity of the scrubbing solution can be calculated from the

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concentrations of the active alkalis present. Note that although HSO_3^- is not basic to dissolved SO_2, it is included in the determination of molar capacity because it is an active alkali but that SO_4^{2-} is not because it is not an active alkali.

"Active sodium" is used to designate the concentration of active alkali in terms of sodium equivalents (Na_2SO_3, $NaHSO_3$, NaOH, etc.). For example, a 1-molar solution of Na_2SO_3 would be 2 molar in terms of active sodium; a 1-molar solution of $NaHSO_3$ would be 1 molar in terms of active sodium. "Inactive sodium" designates the concentration in sodium equivalents of anions that are not active alkalis--this is essentially the concentration of SO_4^{2-} expressed in sodium equivalents. The sodium itself is, of course, a spectator ion that takes no part in any of the reactions.

"Sulfate regeneration" is sometimes used as a synonym for sulfate removal; it has nothing to do with conversion of SO_4^{2-} to active alkali. "Softening" refers to the addition of compounds to reduce the dissolved calcium concentration by precipitating a calcium compound in dilute double-alkali processes. Carbonate softening refers to the addition of a carbonate or CO_2; sulfite softening refers to addition of a sulfite. "TOS"--total oxidizable sulfur--is used to designate the concentration of sulfite (essentially SO_3^{2-} plus HSO_3^-).

There are two types of sodium-based double-alkali processes, differentiated by the concentration of the scrubbing solution. The difference is a result of the sulfate removal method adopted. "Dilute" processes have scrubbing solutions with active alkali concentrations below about 0.15 molar (8). Under these conditions, sulfates can be removed by direct precipitation. The high SO_2 absorption efficiencies attainable with more concentrated solutions are precluded, but for conditions in which high oxidation rates prevail (low-sulfur coal applications or industrial boilers with high excess air rates, for example), dilute processes may be a more desirable alternative than complicated sulfate removal procedures. "Concentrated" processes have scrubbing solutions with active sodium concentrations greater than 0.15 molar, which precludes removal of sulfate by direct precipitation. In some cases, indirect removal of sulfates (coprecipitation with $CaSO_3 \cdot 1/2H_2O$ and losses in the waste) is sufficient to control the sulfate concentration (typically in high-sulfur coal utility applications); in others, the higher efficiency of the more concentrated scrubbing solution may justify an additional sulfate removal process.

Absorption and Regeneration Chemistry

The primary reaction in the absorber consists of the reaction of the sulfite ion with dissolved SO_2.

$$2Na^+ + SO_3^{2-} + SO_2 \cdot H_2O \rightleftarrows 2Na^+ + 2HSO_3^-$$

The role of other anionic species such as CO_3^{2-} and OH^- does not appear to be quantified. Since the SO_3^{2-} concentration is very high compared

with these species, its role in SO_2 absorption is probably minor in most cases, particularly in concentrated double-alkali systems. Since all of the reactants remain in solution, there are no complicating and rate-limiting solid-liquid reactions. Gas-liquid mass transfer becomes more important, if not limiting (9), in the SO_2 absorption mechanism.

The pH of the scrubbing solution remains an important factor, however. Between a pH of 6 and 9, the fraction of the sulfite species that exists as SO_3^{2-} increases from less than 10% to over 90% as the sulfite equilibrium series is driven to the right by the reduction in hydrogen ion concentration

$$SO_2 \cdot H_2O \rightleftarrows HSO_3^- + H^+ \rightleftarrows SO_3^{2-} + 2H^+$$

The molar capacity at equal active alkali concentrations is thus several fold greater at the higher pH. A pH of 8 to 9 appears to be the practical operating limit because carbonate scaling--caused by high concentrations of CO_3^{2-} formed from dissolved CO_2--occurs at higher pHs.

The possible pH operating ranges are a major difference between sodium/lime and sodium/limestone processes. In sodium/lime processes, the pH of the regenerated scrubbing solution can be as high as 12, allowing operation over the entire practical pH range. In sodium/limestone processes, the regenerated scrubbing solution pH cannot be increased over about 7, which is thus the maximum operating pH. Unlike calcium-based slurry processes, however, the alkalinity of sodium-based scrubbing solutions can be increased at a constant pH by increasing the concentration of the dissolved sulfites; process differences between sodium/lime and sodium/limestone processes do not necessarily involve differences in L/G ratios or gas-liquid contact time. Comparing data from pilot-plant studies, for example, Chang and others (10) cite equal SO_2 removal efficiencies for a sodium/lime system with a regenerated scrubbing solution pH of 12, a L/G ratio of 6 to 8 gal/kaft^3, and an active sodium concentration of 0.44 mol/l; and for a sodium/limestone system with a regenerated scrubbing solution pH of 6.6, a L/G ratio of 3 to 4 gal/kaft^3 and an active sodium concentration of 1.35 mols/l.

A portion of the circulating scrubbing solution is withdrawn and treated with lime or limestone to precipitate $CaSO_3 \cdot 1/2H_2O$, simultaneously increasing the solution pH to increase the SO_3^{2-} concentration of the remaining sulfites. In the case of lime the reaction is

$$2Na^+ + 2HSO_3^- + Ca^{2+} + 2OH^- \rightarrow 2Na^+ + SO_3^{2-} + CaSO_3 \cdot 1/2H_2O + 1\text{-}1/2H_2O$$

In the case of limestone, the reaction is

$$2Na^+ + 2HSO_3^- + Ca^{2+} + CO_3^{2-} \rightarrow 2Na^+ + SO_3^{2-} + CaSO_3 \cdot 1/2H_2O + CO_2 + 1/2H_2O$$

Both the lime and limestone reactions are quite efficient. Lime stoichiometries are well demonstrated (4) and can be expected to range from 0.98 to 1.05 mols of $Ca(OH)_2$ per mol of SO_2 absorbed (stoichiometries less than 1.0 are possible because of other alkalis such as makeup Na_2CO_3 that are added to the system). Limestone stoichiometries are not as well defined but stoichiometries of 0.97 have been obtained in pilot tests (10).

Sodium losses in the waste depend on the dewatering efficiency and the sodium concentration in the scrubbing liquid. Sodium makeup rates of 0.05 mol of Na_2CO_3 per mol of SO_2 absorbed are well demonstrated for sodium/lime processes (10). Sodium losses can be expected to be higher in sodium/limestone processes if higher concentration scrubbing solutions are used but the rates have not been well demonstrated.

Oxidation

As in other FGD processes, some of the SO_2 absorbed is oxidized to sulfate by oxygen absorbed from the flue gas. In sodium-based double-alkali processes, this can be a problem of some complexity. In lime and limestone slurry processes, sulfate not removed as a coprecipitate in $CaSO_3 \cdot 1/2H_2O$ can normally be precipitated as gypsum because of the high dissolved calcium (roughly 10^{-5} molar) and low dissolved sulfite (10^{-3} molar or less) concentrations of the liquid. In sodium-based processes, the dissolved calcium concentrations are lower (roughly 10^{-6} molar) and the dissolved sulfite concentrations are higher (sometimes approaching 10^{0} molar). The sulfate cannot be precipitated by adding calcium because of the solubility relationship of calcium sulfate and calcium sulfite. Calcium sulfate is over 100 times as soluble as calcium sulfite. Expressed in terms of solubility product ratio:

$$\frac{Ca^{2+}\ SO_3^{2-}}{Ca^{2+}\ SO_3^{2-}} = \frac{SO_3^{2-}}{SO_3^{2-}} \frac{9.1 \times 10^{-6}}{6.8 \times 10^{-8}} = 134$$

Theoretically (excluding various nonideal deviations), addition of calcium will precipitate only $CaSO_4$ above a SO_4^{2-}/SO_3^{2-} ratio of 134 and only $CaSO_3$ below a ratio of 134. This holds essentially true in practice. Above active-sodium concentrations of about 0.15 molar in double-alkali processes (the range of concentrated processes), sulfates cannot be removed by direct precipitation.

Some sulfate is removed from double-alkali systems by coprecipitation with the $CaSO_3 \cdot H_2O$ as it is in limestone and lime slurry processes (11). Apparently this is sufficient to control sulfate concentrations below oxidation rates of about 20% of the SO_2 absorbed. Some dissolved sulfate is also removed in the residue liquid in the waste. In many utility applications with high-sulfur coal, this means of removal is apparently sufficient (12). Sulfate oxidation in double-alkali processes is usually lower than the rate in limestone and lime slurry processes because the SO_2 absorption rate is enhanced by the high alkalinity of the scrubbing solution, while O_2 absorption, if affected at all, is inhibited by the high ionic strength.

At higher oxidation rates in concentrated double-alkali processes, however, extraordinary means must be taken to remove the sulfate. If allowed to increase, it increases the ionic strength of the scrubbing solution and

decreases the alkalinity (the active sodium concentration) and eventually causes gypsum scaling in the absorber. Sulfate removal in these cases involves a separate process (simple bleedstreams are regarded as environmentally unsound because the waste is soluble). Some of the treatments used are addition of sulfuric acid to a bleedstream to precipitate gypsum and processing of a bleedstream in an electrolytic cell to produce gypsum (7).

SODIUM/LIME DOUBLE-ALKALI PROCESSES

Several U.S. companies have been active in the development of sodium/lime double-alkali FGD technology. These include Arthur D. Little, Inc. (ADL), FMC Corporation, Envirotech Corporation, and General Motors Corporation. Over the past decade, numerous pilot plants and several full-scale industrial double-alkali FGD systems have been operated and five full-scale utility systems are in operation or in the construction or planning stage. The utility systems (13) are shown in Table 1. EPA has been actively involved in double-alkali studies at the EPA Research Triangle Park, Industrial Environmental Research Laboratory (RTP/IERL) in North Carolina and in EPA-sponsored laboratory, prototype, and full-scale evaluations.

TABLE 1. CURRENT SODIUM/LIME DOUBLE-ALKALI UTILITY FGD APPLICATIONS IN THE UNITED STATES

Utility	Station	Unit No.	FGD, MW	Status
Central Illinois Public Service Co.	Newton	1	575	Operating
Louisville Gas and Electric Co.	Cane Run	6	277	Operating
Southern Indiana Gas and Electric Co.	A. B. Brown	1	250	Operating
Northern Indiana Public Service	Schahfer	17	421	Under construction
Northern Indiana Public Service	Schahfer	18	421	Planned

In 1972, ADL was awarded an EPA contract to develop sodium/lime double-alkali technology (8). In the ADL laboratory program, tests were conducted to

develop process chemistry, to study regeneration of sodium-scrubbing solutions, and to characterize double-alkali waste products. ADL, in conjunction with Combustion Equipment Associates, Inc. (CEA), conducted pilot-plant work involving both concentrated and dilute modes of operation. A 20-MW prototype concentrated mode system was designed by ADL and CEA and evaluated at Gulf Power Company's Scholz Steam Plant at Sneads, Florida (14). The test program, a part of the EPA contract, ran from May 1975 to July 1976. ADL and CEA were awarded a contract for a full-scale concentrated mode unit now operating on unit 6 of Louisville Gas and Electric Company's Cane Run Station (15). The sulfur content of the coal used in unit 6 ranges from 3.5% to 4.0%. The cost of the installation was borne by Louisville Gas and Electric and EPA provided funding for performance testing and a 1-year operational study.

After working with wet-limestone and lime scrubbing, FMC began developing a sodium-based scrubbing process on a pilot-plant scale in the 1960s. The process was intended to produce sodium sulfite and sodium sulfate for marketing. When efforts to sell these products failed, FMC began development of a method of recovering the sodium from the system while producing an acceptable solid waste for disposal. The resulting concentrated mode process was demonstrated as a 30-MW equivalent prototype system on a reduction kiln at FMC's Modesto (California) Chemical Plant in 1971 (16). Since 1971, several installations have been constructed for industrial boilers of up to 150-MW equivalent size. A FMC system, scrubbing flue gas from a 3.75% sulfur coal, is operating on the 250-MW unit 1 at Southern Indiana Gas and Electric Company's A. B. Brown Station (17). In addition, FMC has a 421-MW system under construction on unit 17 at Northern Indiana Public Service's Schahfer station and another 421-MW system is being planned for unit 18.

Envirotech Corporation developed its Buell double-alkali SO_2 control process for both dilute and concentrated mode operation (18). A joint development effort with Utah Power and Light Company was undertaken with a 1-MW pilot plant at the Gadsby Station in Salt Lake City, which was started up in January 1972. Research by Envirotech has also focused on high-chloride coals and the acceptable disposal of chlorides in a waste-producing system. Envirotech constructed a 575-MW FGD system on unit 1 at Central Illinois Public Service Company's Newton Station. Unit 1, which burns a 4% sulfur coal, was started up in 1979.

SODIUM/LIMESTONE DOUBLE-ALKALI PROCESS

With the maturation of sodium/lime double-alkali technology and its movement into commercial application, EPA work in double-alkali FGD has shifted to development of sodium/limestone processes. The reasons for using limestone are primarily economic; limestone costs approximately 8 $/ton while lime costs 75 to 100 $/ton, depending on the delivery costs. The major drawback of limestone is that it is less reactive than lime and the regeneration reaction tanks must provide a longer residence time.

Much of the EPA-sponsored development work has been done on a small bench-scale unit by IERL at RTP. However, during 1980 and early 1981, EPA and

others funded a 20-MW prototype demonstration of the sodium/limestone double-alkali process (as developed by Thyssen-CEA Environmental Systems, Inc., and ADL) at Gulf Power Company's Scholz Station (10). This test program used existing equipment from the prototype demonstration of the ADL-CEA sodium/lime double-alkali technology during the mid-1970s.

Modification and conversion of the existing system began in early 1980 and initial operations began in August. Formal testing of the process did not start until February 1981 and then only lasted two months because of financial constraints compounded by mechanical problems with the existing equipment. Thus, although some data were obtained, the quantity and quality of the data were limited. For this reason, IERL is continuing development work with their bench-scale unit at RTP.

Although the test program was significantly shorter than originally anticipated, some meaningful test results were obtained and have been published. The absorption area seemed to perform as expected, with SO_2 removal efficiency averaging 96%. The inlet SO_2 concentration varied from 1,460 to 3,240 ppm. The limestone utilization (as determined from chemical analyses of the filter cake) ranged from 93% to 100% and averaged over 97%.

However, the problems cast a shadow over some of the technical results. Because the pilot plant did not achieve a long-term, closed-loop operation, several process assumptions could not be confirmed. For example, the effectiveness of sulfate removal by precipitation with calcium sulfite in the regeneration system was not confirmed. Although the sulfate concentration in the system remained at an acceptable steady-state level during the test program, liquid losses contributed to the removal of sulfate from the system. Additional test work is needed to confirm that acceptable sulfate levels can be maintained in a closed-loop operation under a variety of operating conditions. In addition, soda ash consumption was excessively high (0.29 mol of $NaCO_3$ per mol of SO_2 removed) because of the failure to achieve closed-loop operation. Theoretically, in a well-designed sodium/limestone double-alkali system, operating in a closed-loop mode, neither the sulfate removal nor the sodium losses are expected to be a significant problem.

Perhaps the most significant problem encountered at Scholz was the poor settling characteristics of the waste solids from the limestone regeneration area. The cause remains unknown and no clear-cut solution for this problem has been determined. In some cases, process changes alleviated the problem, while at other times, the settling characteristics deteriorated until the system had to be shut down. The solids content in the filter cake was typically 35% to 45%, much below the expected (and desired) 55%. Additional development work is needed to determine the cause of, and the solution to, the solids settling characteristics and the effects of the solution on the design of the sodium/limestone double-alkali process.

DESIGN AND ECONOMIC PREMISES

The premises used in this study have been developed by TVA, EPA, and others during similar economic evaluations made since 1967. The design premises are formulated to quantify SO_2 emissions from a representative electric utility coal-fired power plant boiler and to establish efficiencies, process flow rates, and other design and operating conditions to meet environmental regulations. The economic premises are designed to represent the many factors affecting costs. They prescribe procedures for determining capital investment and annual revenue requirements based on regulated utility economics. Results are reported in sufficient detail to permit comparisons with previous economics and to permit adjustments based on use of alternative premises. The costs are based on 1982 costs for capital investment and 1984 costs for annual revenue requirements.

The economic evaluations are based on a conceptual design developed from the design premises and engineering data such as flow diagrams, material balances, and equipment costs. The base case is a new 500-MW power unit burning 3.5% sulfur, 16% ash bituminous coal, and complying with the 1979 NSPS (19).

DESIGN PREMISES

Coal Premises

The coal used is an eastern bituminous coal containing 3.5% sulfur on a dry basis. The properties of the coal are based on composites of many samples representing major coal production areas (20,21). The coal composition is shown in Table 2.

Power Plant

The power plant site is in the north-central region (Illinois, Indiana, Ohio, Michigan, and Wisconsin). The location represents an area in which coal-fired power plants burning coals of diverse type and source are situated (22). The design is based on standard design practices (23,24) and current trends in utility boiler construction (25).

The base case power unit is a new, single 500-MW, balanced-draft, pulverized coal-fired, dry-bottom boiler. The heat rate is 9,500 Btu/kWh. To provide for equitable comparisons, the power unit is not derated for energy consumption by the FGD systems evaluated. Instead, the energy requirements are charged as independently purchased commodities. Cost estimates are based

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on a single power unit independent of other units at the site. The FGD system being evaluated is assumed to be installed during construction of the power plant and to have a 30-year life and to operate full load for 5,500 hours a year. This operating schedule is equivalent to a total lifetime operation of 165,000 hours.

TABLE 2. PREMISE COAL COMPOSITION

Coal	Sulfur, %	Ash, %	Moisture, %	Heat content, Btu/lb	Ultimate analysis, % C.	H.	O.	N.	Cl
Eastern bituminous As-fired basis:	3.36	15.14	4.0	11,700	66.7	3.8	5.6	1.3	0.1
Moisture-free basis:	3.50	15.7	-	-	69.5	4.0	5.8	1.4	0.1

Flue Gas Composition

The flue gas composition is based on combustion of pulverized coal, assuming a total air rate equivalent to 139% of the stoichiometric requirement (the air required for combustion of carbon, hydrogen, and sulfur). This includes 20% excess air to the boiler and 19% additional air leakage to the flue gas in the air heater. It is assumed that 80% of the ash present in each coal is emitted as fly ash and 95% of the sulfur is emitted as SO_x. The remaining sulfur is removed in the bottom ash and fly ash. No loss of sulfur in the pulverizers is assumed. Three percent of the sulfur emitted is SO_3 and the remainder is SO_2. The flue gas composition and flow rates are shown in Table 3.

Environmental Regulations

Emissions from new coal-fired utility plants are regulated by NSPS and promulgated by EPA under authority of Section 111 of the Clean Air Act (Public Law 91-604). This section requires EPA to set Federal emission limitations that reflect the degree of control that can be achieved by using the best technological system of continuous emission reduction which has been adequately demonstrated. In 1979, in response to the Clean Air Act Amendments of 1977 (Public Law 95-95), EPA promulgated revised NSPS for emissions of SO_2, NO_x, and particulate matter from utility power plants upon which construction began or begins after September 1978 (19). These 1979

NSPS, shown in Table 4, are used as the FGD design basis in this study. The controlled outlet SO_2 emission and SO_2 removal efficiency required are shown in Table 5.

TABLE 3. FLUE GAS COMPOSITION

Component	Volume, %	Lb-mol/hr	Lb/hr
N_2	75.18	137,718	3,856,117
O_2	5.53	10,134	324,300
CO_2	12.31	22,552	992,299
SO_2	0.21	392	25,113
SO_3	0.01	12	970
NO	0.03	59	1,766
NO_2	0.00	3	142
HCl	0.01	12	418
H_2O	6.72	12,287	221,163
	100.00	183,169	5,422,288
Fly ash			50,046
Total			5,472,334

Sft3/min (60°F) = 1,156,000 to FGD system
Aft3/min (300°F) = 1,690,000 to FGD system

Fly ash loading

	Gr/sft3
Wet	5.05
Dry	5.41

Sulfuric acid dewpoint temperature: 308°F

Basis: 500-MW power unit, 3.5% sulfur coal.

Particulate Matter

Cold-side (after the air heater) electrostatic precipitators (ESP) sized to meet the 0.03 lb/MBtu standard are assumed for particulate matter control. The costs for ash collection and disposal are not included in the economic

evaluations. Although fly ash is used in the sodium/limestone double-alkali process to fix the waste, the costs of collection and handling are identical for both processes. Costs are calculated for the entire sodium/limestone double-alkali process waste disposal area and a credit is given for the portion of the waste disposal area occupied by fly ash.

TABLE 4. 1979 NSPS EMISSION STANDARDS

SO_2

70% SO_2 removal (minimum) to a maximum SO_2 emission of 0.6 lb SO_2/MBtu
0.6 lb SO_2/MBtu maximum emission up to 90% SO_2 removal
90% SO_2 removal (minimum) to a maximum SO_2 emission of 1.2 lb SO_2/MBtu
1.2 lb SO_2/MBtu maximum emission

NO_x (not considered in this study)

Bituminous coal - 0.6 equivalent lb NO_2/MBtu
Subbituminous coal - 0.5 equivalent lb NO_2/MBtu
Lignite - 0.6 equivalent lb NO_2/MBtu

Particulate

0.03 lb/10^6 Btu

TABLE 5. EMISSION STANDARD AND REQUIRED SO_2 REMOVAL EFFICIENCY

Coal	Equivalent SO_2 content of coal, lb SO_2/MBtu	Overall equivalent SO_2 removal efficiency, %	Equivalent SO_2 removal required in FGD system, %[a]	Controlled outlet emission, lb SO_2/MBtu
Eastern bituminous, 3.5% S	5.74	89.6	88.7	0.60

a. Based on FGD as the only SO_2 control device and 5% sulfur retention in the ash.

Flue Gas Desulfurization

The FGD systems are coupled to the power unit by a plenum into which the power plant induced draft (ID) fans discharge. The plenum allows the scrubbing systems to be designed for a different number of trains than the number of power plant ducts (to account for limitations in the available size of absorber units), and it facilitates the use of spare absorber trains. To minimize flow control problems that can result from this design, separate ID booster fans are provided for each train of the FGD system downstream from the reheater to overcome the pressure drop in the FGD train.

Absorber trains are sized for a maximum of 125 MW of flue gas up to a maximum of 513,000 sft^3/min (60°F). Thus for full scrubbing, the 500-MW base case requires four 125-MW operating trains. It is assumed that the annual availability of an absorber train is 85% and that no scrubbing time is lost during startup. A spare absorber train is provided to allow use of an emergency bypass. The FGD system is designed for closed-loop operation in which the only liquid water loss is in the waste. This loss is assumed to be sufficient for chloride control. Material balances are determined on this basis.

Emergency Bypass--

Because the 1979 NSPS allow emergency bypass around the FGD system under some conditions if spare scrubbing capacity is provided, a spare absorber train and provision for bypass of 50% of the scrubbed gas are included. An emergency bypass of 50% of the scrubbed gas is assumed to be an economic balance between the higher cost of providing additional bypass and the small likelihood of multiple absorber train failures making higher bypass rates necessary. The bypass is installed as two identical ducts from each end of the inlet plenum to the stack plenum.

Ductwork--

Square ductwork with 2-inch insulation is used for the inlet plenum and absorber trains. A gas velocity of 50 ft/sec is used for the inlet plenum, all ductwork, and the emergency bypass. A gas velocity of 25 ft/sec is used for the reheater section. Duct material is 3/16-inch Cor-Ten steel when the gas temperature is higher than 150°F and 3/16-inch stainless steel when the gas temperature is lower than 150°F.

Removal Efficiencies--

It is assumed that 50% of the SO_3, 95% of the HCl, none of the NO_x, and 50% of the remaining fly ash in the flue gas are removed in the FGD system. It is assumed that 5% of the SO_2 is removed in the presaturator and that the remaining SO_2 removal takes place in the absorber.

Spare Equipment--

Spare equipment consists of a spare train of crushing and grinding equipment in the limestone preparation area, a spare absorber train, a spare waste filter, and spare process pumps in each train.

Mist Eliminator--

The mist eliminator is a 3-pass, open-vane, chevron type. The mist eliminator reduces entrained moisture to a maximum level of 0.1% (by weight) of the flue gas. This maximum level is assumed for calculation of the amount of stack gas reheat required.

Flue Gas Reheat--

Indirect steam reheat is provided to evaporate entrained water droplets not removed by the mist eliminator and to maintain a 175°F temperature at the stack. Upstream reheater tubes in contact with gas below 150°F are Inconel 625; the remainder are Cor-Ten steel. The reheater raises the temperature of the gas at the exit to approximately 170°F. The remaining 5°F of reheat is provided by the heat of compression in the booster fans downstream from the reheaters.

Raw Materials--

Storage capacity for a 30-day supply of raw materials is provided. Standard raw material characteristics are shown in Table 6.

TABLE 6. RAW MATERIAL CHARACTERISTICS

Size as received	Ground size	Analysis[a]	Bulk density, lb/ft^3
Limestone 0 x 1-1/2 inches	90% to pass 325 mesh	95% $CaCO_3$ 0.15% MgO 4.85% inerts 5 lb H_2O/100 lb dry limestone	95
Fineness of grind index factor = 5.7 Hardness of work index factor = 10			
Soda ash: 100% to pass 100 mesh	-	99.8% Na_2CO_3 0.15% NaCl 0.02% inerts 0.03% H_2O	35
Lime (agricultural)	-	95.0% CaO 0.15% MgO 4.85% inert	55

a. Dry basis.

Waste Disposal

The wastes generated by the FGD processes are disposed of in a landfill one mile from the FGD facilities. Sufficient land is provided for disposal for the life of the FGD facility. The disposal site is assumed to be an area of low relief with sufficient soil for landfill requirements.

A square landfill with a 20-foot-high perimeter and a 6-degree cap is used, as shown in Figure 1. After topsoil removal, the landfill area is lined with 12 inches of clay (assumed available onsite). A French drain system (perforated pipe surrounded by gravel) and 24 inches of bottom ash are laid on top of the clay. The bottom ash layer and French drain system allow the seepage to be collected separately from runoff and treated for pH adjustment.

Land requirements include the landfill, catchment basin, equipment storage area, topsoil storage area, and a 50-foot perimeter of undisturbed land. Costs for access roads; a 6-foot security fence around the landfill area; security lighting; and topsoil stripping, replacement, and revegetation are included. One upstream and three downstream groundwater monitoring wells are also included.

ECONOMIC PREMISES

A 3-year construction period, from early 1981 to late 1983, is used. Mid-1982 costs are used for capital investment. Mid-1984 costs are used for annual revenue requirements. These costs represent the midpoint of construction expenditures in 1982 and the midpoint of the first year of operation in 1984. Costs are projected from Chemical Engineering cost indexes, as shown in Table 7, using standard estimating techniques. Frequently used costs are shown in Table 8.

TABLE 7. COST INDEXES AND PROJECTIONS

Year	1978	1979	1980[a]	1981[a]	1982[a]	1983[a]	1984[a]
Plant	218.8	238.7	257.8	277.1	297.9	320.2	342.6
Material[b]	240.6	264.4	288.2	311.2	336.1	363.0	388.4
Labor[c]	185.9	194.9	210.5	227.3	245.5	265.2	283.7

a. TVA projections.
b. Same as "equipment, machinery, supports" Chemical Engineering index.
c. Same as "construction labor" Chemical Engineering index.

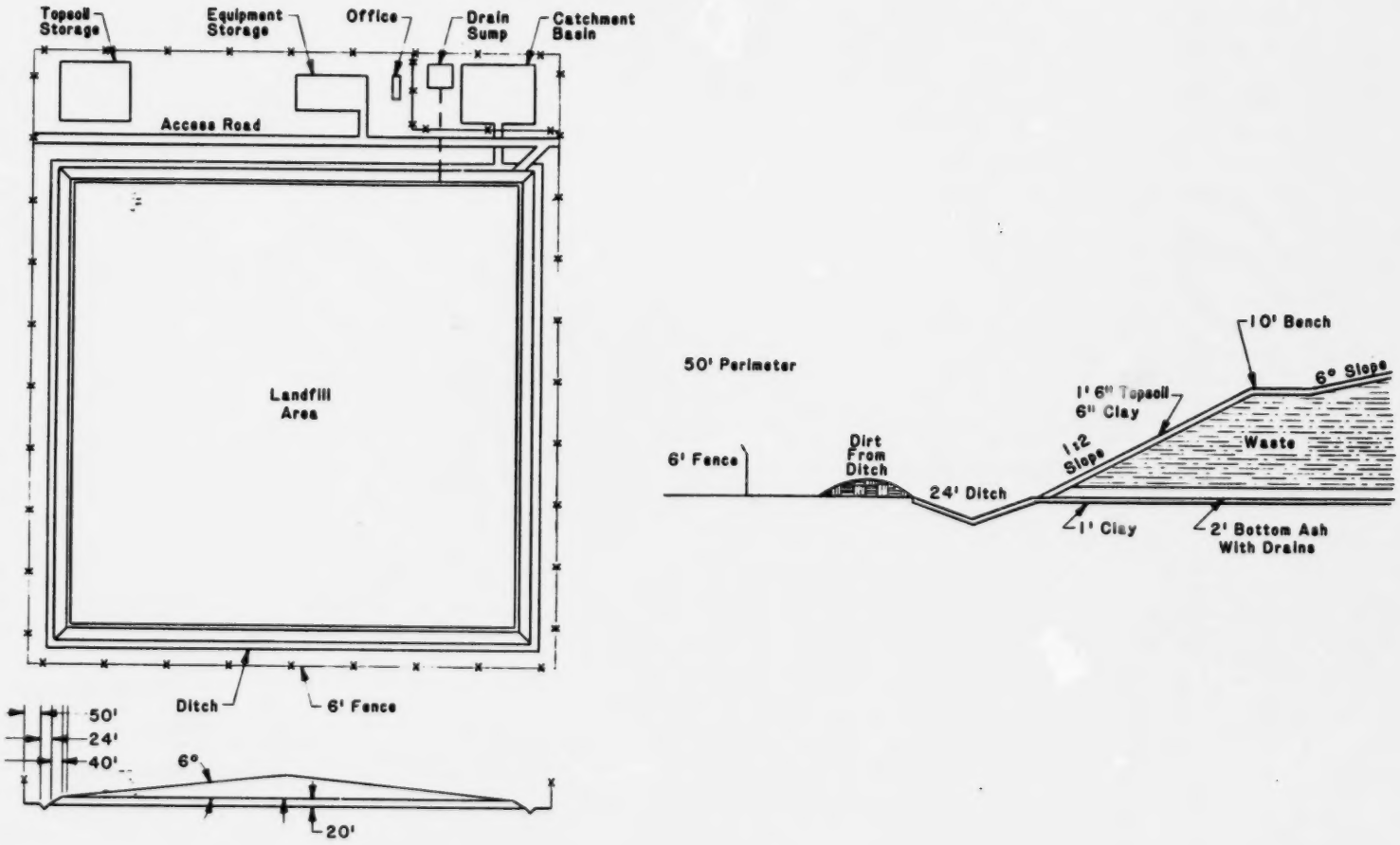

Figure 1. Landfill layout.

TABLE 8. COST FACTORS

1984 Utility Costs	
Electricity	0.037 $/kWh
Steam	2.50 $/klb, 3.30 $/MBtu
Diesel fuel[a]	1.60 $/gal
Filtered river water	0.14 $/kgal
1984 Labor Costs	
FGD operating labor	15.00 $/man-hr
Waste disposal labor	21.00 $/man-hr
Analyses labor	21.00 $/man-hr
1984 Raw Material Costs	
Limestone	8.50 $/ton
Soda ash	160 $/ton
Lime	75 $/ton
1982 Land Cost	5,000 $/acre

These cost factors are based on a north-central plant location.

a. Cost is based on wholesale price of barge-load quantities. Road taxes are not included.

Capital Cost Estimates

The capital investment estimates are divided into three major sections: direct investment, indirect investment, and other capital investment. The sum of direct and indirect investments is the fixed investment.

Direct Investment--

Direct investment consists of the installed costs of all process equipment, including provisions for services, utilities, and miscellaneous and waste disposal investment. Installation costs are estimated by major processing area and include charges for all piping, foundations, excavations, structural steel, electrical equipment, instruments, ductwork (all flue gas ductwork is included in the gas-handling area), paint, buildings, taxes, freight, and a premium for 7% overtime construction labor.

Service facilities are allocated on the basis of process requirements. Included in the utilities investment are necessary electrical substations, process water, fire and service water, instrument air, chilled water, inert gas, and compressed air facilities. Services, utilities, and miscellaneous are estimated as 6% of the total process capital.

All equipment and direct construction costs associated with waste disposal are included in the landfill costs. All mobile equipment involved in loading and transporting the waste from the in-process storage area, as well as working the landfill, is included in landfill equipment. The sum of total process capital and the waste disposal cost is the total direct investment.

Indirect Investment--

Indirect investment comprises fees for engineering design and supervision, architect and engineering contractor, construction expense, contractor fees, and contingency. The percentages used to calculate these costs are shown in Table 9.

TABLE 9. INDIRECT CAPITAL COST FACTORS

	Indirect investment, % of direct investment	
	Process	Landfill
Engineering design and supervision	7	2
Architectural and engineering contractor	2	1
Construction expense	16	8
Contractor fees	5	5
	30	16

The contingency fee is a percentage of the direct and other indirect investments based on the level of technical development. A 20% contingency is used for the sodium/limestone double-alkali process and a 10% contingency is used for the forced-oxidation limestone process excluding landfill. A 20% contingency is used for the landfill in both processes.

Other Capital Investment--

The allowance for startup and modifications is applied as a percentage of the total fixed investment. Since the startup and modification costs for the landfill are assumed to be negligible, this allowance is calculated using only the process fixed investment. A 10% startup charge is applied to the sodium/limestone double-alkali process because of its developmental status while 8% is charged for the forced-oxidation limestone process.

The cost of borrowed funds (interest) during construction is 15.6% of the fixed investment (both process and landfill). This factor is based on an assumed 3-year construction schedule and is calculated with a 10% weighted cost of capital, with 25% of the construction expenditures in the first year, 50% in the second year, and 25% in the third year of the project. Expenditures in a given year are assumed uniform over that year. Startup costs are assumed to occur late in the project schedule so that there are no charges for the use of money to pay startup costs. No royalty charges are included for the forced-oxidation limestone process because of its generic design, but a 1% charge is included for the sodium/limestone double-alkali process.

Land--

All land associated with the process (including the landfill) is charged to the process. The cost of land is $5,000 per acre.

Working Capital--

Working capital is the amount of money invested in raw materials, supplies, accounts receivable, and money on deposit for payment of operating expenses. Working capital is the equivalent cost of one month's raw material cost, 1.5 months' conversion cost, 1.5 months' plant and administrative overhead costs, and 3% of the total direct capital investment. One month is defined as 1/12 of annual costs.

Annual Revenue Requirements

Annual revenue requirements account for recovery of direct and indirect operating and maintenance costs and capital charges. Annual revenue requirements normally vary from year to year as operating and maintenance costs change and capital charges decline. Thus, no single year is necessarily representative of the lifetime costs, and single-year undistorted comparisons cannot be made among processes with different ratios of operating costs to capital charges. In addition, it is necessary to take into account the effect of time on the value of money (i.e., for inflation, the future earning power of money spent, and other factors).

Frequently these factors are accounted for by levelizing (26). Levelization converts all the varying annual revenue requirements to a constant annual value, such that the sum of the present worths of the levelized annual revenue requirements equals the sum of the present worths of the actual annual revenue requirements. The levelized value is calculated by multiplying the revenue requirements for each year by the appropriate present worth factor and summing the present worth values. Then the single present worth value is converted to equal annual values by multiplying the result by the capital recovery factor.

In this study, the operating and maintenance costs are levelized by multiplying the first-year operating and maintenance costs by a levelizing factor. The levelized capital charges are determined by levelizing the percentage of capital investment applied yearly as capital charges. The levelizing factor includes a discount factor reflecting the time value of money and an inflation factor reflecting the effects of inflation during the

operating life of the system. The discount rate used is 10% and the inflation rate used is 6%.

Operating and Maintenance Costs--

Operating and maintenance costs consist of direct costs for raw materials and conversion and indirect costs for overheads. Conversion costs consist of operating labor and supervision, utilities, maintenance, and analysis costs.

In this evaluation, the raw materials consist of limestone, lime, and soda ash. Raw material costs are determined from vendor quotations or published sources. All costs are delivered costs.

Operating labor and supervision consists of all labor requirements for operation of the FGD-related equipment and the landfill. The allocation of operating labor and supervision man-hours for the FGD system depends on the process complexity, number of process areas, labor intensity of the process, and operating experience. Waste disposal operating labor and supervision depends on the amount of equipment required to operate the landfill.

Costs for steam and electricity are based on the assumption that they are purchased from another source; this eliminates the need to derate the power unit. Electrical power requirements are determined from the installed horsepower of operating electrical equipment (excluding the horsepower of spare equipment). Each motor in operation is assumed to be operating at rated capacity, although this results in higher power consumptions than would actually occur. Diesel fuel is calculated based on the equipment required for the landfill.

Process maintenance costs are calculated as a percentage of direct process investment, which varies depending on the process as shown in Table 10. Landfill maintenance costs are 3% of the landfill direct investment (equipment and construction). Analysis costs are based on the complexity of the process and are listed as a single entry.

TABLE 10. MAINTENANCE FACTORS

	Maintenance, % of total direct investment	
	FGD system	Waste disposal
Limestone scrubbing	8	3
Double-alkali	6	3

Plant and administrative overheads include plant services, general engineering (excluding maintenance), transportation, and the expenses connected with management activities. Plant and administrative overheads are 60% of the conversion costs less utilities.

Capital Charges--

Capital charges are those costs incurred by construction of the facility that must be recovered during its life. They consist of returns on equity and debt (discount rate), depreciation, income taxes, and other costs such as insurance and local taxes. In keeping with common practice for investor-owned utilities, the weighted cost of capital is used as the discount rate (27). Depreciation is expressed as a sinking fund factor to simplify calculations. An allowance for interim replacement is included. Credits are also included for tax preference allowances. The capital charges are shown below. Levelized capital charges are used for both first-year and levelized annual revenue requirements.

	% of total capital investment
Weighted cost of capital	10.00
Depreciation (sinking fund factor)	0.61
Annual interim replacement	0.56
Levelized accelerated tax depreciation	-1.36
Levelized investment tax credit	-1.93
Levelized income tax	4.31
Insurance and property taxes	2.50
Levelized annual capital charges	14.7

The capital charges are applied as a percentage of the total capital investment, including land and working capital. Although land and most working capital cannot be depreciated and are not subject to investment tax credit, their inclusion simplifies the calculation and has an insignificant effect on capital charges.

The capital structure is assumed to be 35% common stock, 15% preferred stock, and 50% long-term debt. The cost of capital is assumed to be 11.4% for common stock, 10.0% for preferred stock, and 9.0% for long-term debt. The weighted cost of capital is 10.0%. The discount rate is equal to the weighted cost of capital. Other economic factors used in financial calculations are a 10% investment tax credit rate, 50% State plus Federal income taxes, 2.5% property tax and insurance, and an annual inflation rate of 6%. Salvage value is assumed to be less than 10% and equal to removal cost.

The sinking fund factor method of depreciation is used since it is equivalent to straight-line depreciation levelized for the economic life of the facility using the weighted cost of capital. The use of the sinking fund factor does not suggest that regulated utilities commonly use sinking fund

depreciation. An annual interim replacement (retirement dispersion) allowance (28) is also included as an adjustment to the depreciation account to ensure that the initial investment will be recovered within the actual rather than the forecasted life of the facility. Tax preference allowances are incentives designed to encourage investment as a stimulus to the overall economy. The basic accounting method used is the flowthrough method which passes the tax advantages to revenue requirements as soon as they occur.

Levelized Operating and Maintenance Costs--

Assuming a constant inflation rate, the levelized operating and maintenance costs are determined by multiplying the first-year operating and maintenance costs by an appropriate levelizing factor, L_f. The levelizing factor is calculated as follows:

$$L_f = CRF_e\,(K + K^2 + K^3 + \text{---} + K^N)$$
$$= CRF_e\,[K(1 - K^N)]/(1 - K)$$

where: CRF_e = capital recovery factor for the economic life
K = $(1 + i)/(1 + r)$; present worth of an inflationary value
i = inflation rate
r = discount rate
N = book life in years

An inflation rate of 6% ($i = 0.06$) and a discount rate of 10% ($r = 0.10$) are used. The first-year operating and maintenance costs are multiplied by the levelizing factor to obtain the levelized operating and maintenance costs.

SYSTEMS ESTIMATED

The sodium/limestone double-alkali FGD process is based on a new, 500-MW coal-fired power unit burning 3.5% sulfur eastern bituminous coal. An equivalent forced-oxidation limestone FGD process, which was available from a previous evaluation (29), is included for comparison. The major design conditions are shown in Table 11.

TABLE 11. PROCESS DESIGN CONDITIONS

	Process	
	Sodium/limestone double alkali	Limestone scrubbing
Absorber type	Sieve tray tower	Spray tower
Superficial gas velocity, ft/sec	9	10
L/G, gal/kaft^3		
Presaturator/underspray	2/1	4/0
Absorber	5	106
Stoichiometry, mol Ca/mol (SO_2 + 2HCl) absorbed	1.0	1.4
Soda ash feed rate, mol Na^+/mol SO_2 absorbed	0.07	-
Sulfite oxidation, %	10	95
Thickener feed solids, %	1.4	8
Thickener underflow solids, %	25	40
Filter cake solids, %	55	85

SODIUM/LIMESTONE DOUBLE-ALKALI PROCESS

The design of the sodium/limestone double-alkali process is based on the previously discussed design premises and the results of the EPA evaluations discussed in the background section. A flowsheet and material balance are shown in Figure 2 and Table 12.

Process Description and Principles of Operation

The FGD plenum distributes the flue gas from the power plant ID fans to the four operating absorber trains. The absorbers are sieve tray towers equipped with presaturators. The flue gas enters the presaturator and is sprayed with recirculating scrubbing solution that humidifies and cools the gas to 125°F. The cooled gas then enters the base of the absorber and rises passing through two sieve trays. The lower sieve tray is sprayed from below with recirculated scrubbing solution. The upper sieve tray is sprayed with regenerated scrubbing solution from above. The scrubbed gas leaves the absorber and passes through a horizontal, chevron-type mist eliminator. The saturated flue gas is reheated to 170°F by an inline steam reheater and passes through the booster fan which heats the gas to 175°F by compression before it enters the stack.

Scrubbing solution from the absorber drains to the recirculating tank. Some of the solution is pumped to the presaturator to supply the cooling sprays and some is pumped to the absorber to wash the underside of the bottom tray. The remaining solution is pumped to the regeneration area where it is treated with fresh limestone slurry in a series of four reaction tanks. The effluent from the last reaction tank is pumped into the thickener which concentrates the solids from 1.4% to 25%. The thickener overflows to a recycle tank into which fresh soda ash solution is added. The resulting regenerated solution is pumped to the top tray in the absorber. The scrubbing solution contains 1.7 mol percent sodium, 0.6 mol percent sulfate, and 0.7 mol percent active alkali (SO_3^{2-} and HSO_3^-) at the absorber inlet and 0.5 mol percent active alkali at the absorber outlet.

The thickener underflow is pumped to the filter feed tank and then pumped to two rotary vacuum filters. The filter cake containing 55% solids is washed to reduce sodium losses. The filter cake is mixed in a pugmill with dry fly ash and a small proportion of lime. The resulting waste material is trucked to an onsite landfill for disposal. The filtrate and wash water are pumped to a surge tank. From this surge tank, a portion is pumped to the ball mill in the limestone preparation area while the remainder is recycled to the thickener.

Soda ash is received by truck or rail, wetted, and stored as a Na_2CO_3 slurry. A saturated solution is withdrawn as needed and pumped to the thickener overflow tank to replace sodium losses from the FGD system. Limestone is received by truck or rail and placed in a storage pile. The limestone is transferred from the storage pile to feed bins. The feed bins feed the gyratory crusher, which in turn feeds the wet ball mill. Filtrate from the

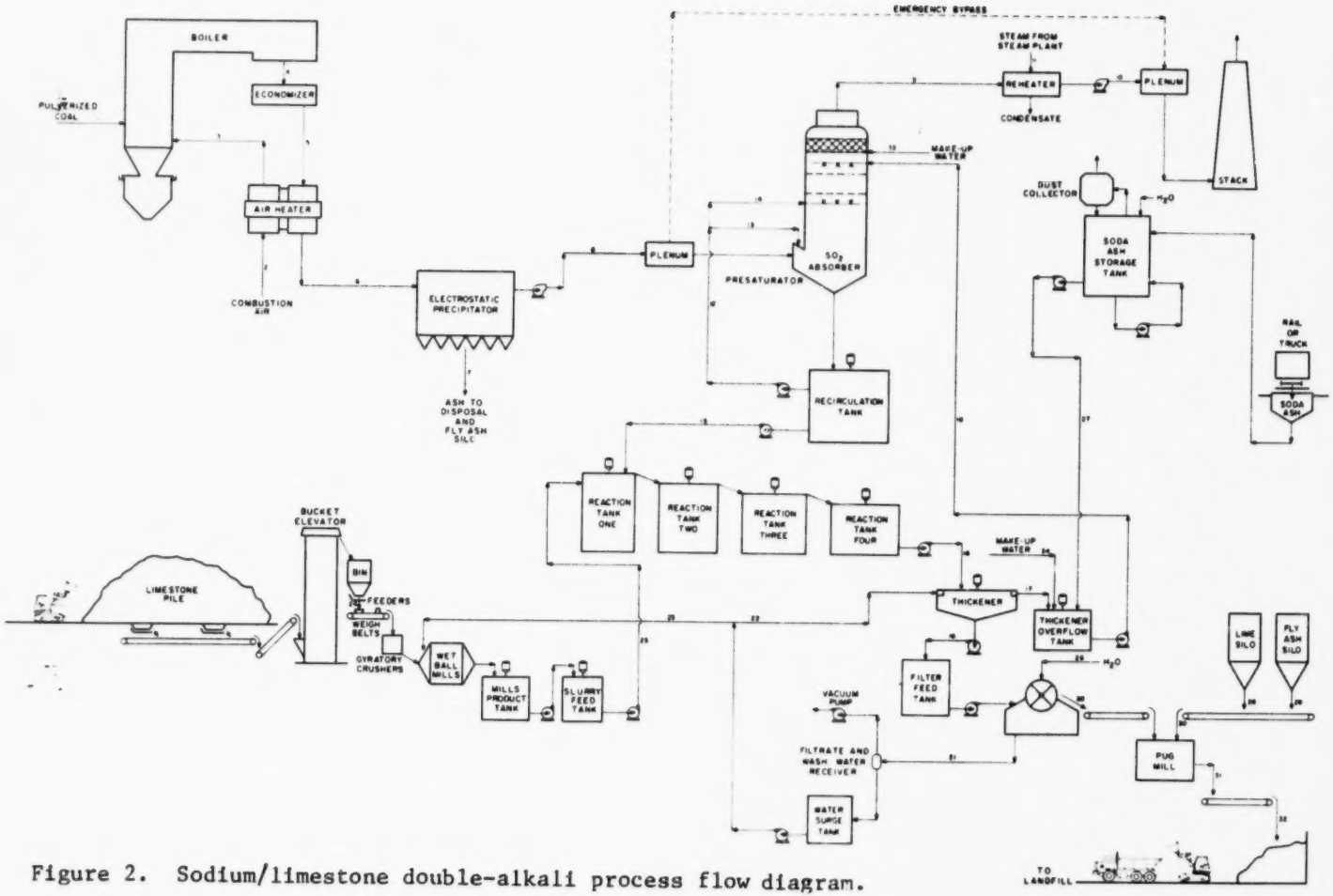

Figure 2. Sodium/limestone double-alkali process flow diagram.

TABLE 12. SODIUM/LIMESTONE DOUBLE-ALKALI PROCESS

MATERIAL BALANCE

	Stream No.	1	2	3	4	5
	Description	Coal to boiler	Total air to air heater	Combustion air to boiler	Gas to economizer	Gas to air heater
1	Total stream, lb/hr	405,800	5,079,000	4,385,000	4,778,000	4,778,000
2	Flow rate, aft3/min(600F)		1,120,000	967,300	1,004,000	1,004,000
3	Temperature, oF		80	495	890	705
4	Pressure, psig					
5	Flow rate, gal/min					
6	Specific gravity					
7	pH					
8	Undissolved solids, %					
9						
10						

	Stream No.	6	7	8	9	10
	Description	Gas to electrostatic precipitator	Fly ash to disposal	Gas to SO_2 absorber	Gas to reheater	Gas to stack
1	Total stream, lb/hr	5,472,000	49,900	5,422,000	5,634,000	5,634,000
2	Flow rate, aft3/min(600F)	1,157,000		1,156,000	1,238,000	1,238,000
3	Temperature, oF	300		300	124	175
4	Pressure, psig					
5	Flow rate, gal/min					
6	Specific gravity					
7	pH					
8	Undissolved solids, %					
9						
10						

	Stream No.	11	12	13	14	15
	Description	Steam to reheater	Recycle to presaturator	Presaturator spray	Underwash spray	Absorber bleed
1	Total stream, lb/hr	96,200	2,148,000	1,432,000	716,100	3,431,000
2	Flow rate, aft3/min(600F)					
3	Temperature, oF	470				
4	Pressure, psig	500				
5	Flow rate, gal/min		4,167	2,778	1,389	6,661
6	Specific gravity		1.04	1.04	1.04	1.04
7	pH					
8	Undissolved solids, %		0	0	0	0
9						
10						

	Stream No.	16	17	18	19	20
	Description	Reactor effluent	Thickener overflow	Regenerated scrubbing solution	Thickener underflow	Filter cake
1	Total stream, lb/hr	3,478,000	3,476,000	3,546,000	192,300	87,400
2	Flow rate, aft3/min(600F)					
3	Temperature, oF					
4	Pressure, psig					
5	Flow rate, gal/min	6,667		6,705	6,945	307
6	Specific gravity	1.05		1.04	1.03	1.24
7	pH					
8	Undissolved solids, %	1.4	0	0	25	55
9						
10						

(Continued)

TABLE 12. (Continued)

	Stream No.	21	22	23	24	25
	Description	Filtrate and wash water	Clear liquid to thickener	Clear liquid to ball mill	Limestone to weigh feeder	Limestone slurry to reactor
1	Total stream, lb/hr	214,100	191,200	22,900	39,100	62,000
2	Flow rate, aft3/min(600F)					
3	Temperature, oF					
4	Pressure, psig					
5	Flow rate, gal/min	422	377	45		77
6	Specific gravity	1.0	1.0	1.0		1.6
7	pH					
8	Undissolved solids, %					60
9						
10						

	Stream No.	26	27	28	29	30
	Description	Filter cake wash water	Fresh soda ash solution	Lime to pug mill	Fly ash to pug mill[a]	Lime and fly ash to pug mill
1	Total stream, lb/hr	169,200	7,000	3,800	48,100	51,900
2	Flow rate, aft3/min(600F)					
3	Temperature, oF					
4	Pressure, psig					
5	Flow rate, gal/min	218	14			
6	Specific gravity	1.0	1.33			
7	pH					
8	Undissolved solids, %					
9						
10						

	Stream No.	31	32	33	34	
	Description	Pug mill product	To landfill disposal	Makeup water to mist eliminator	Makeup water to thickener overflow tank	
1	Total stream, lb/hr	139,400	139,400	100,100	62,300	
2	Flow rate, aft3/min(600F)					
3	Temperature, oF					
4	Pressure, psig					
5	Flow rate, gal/min			200	124	
6	Specific gravity	1.36	1.36	1.0	1.0	
7	pH					
8	Undissolved solids, %					
9						
10						

1						
2						
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a. Remaining fly ash disposed with bottom ash.

solids separation area is added to the ball mill to produce a 60% slurry, which is stored in the mills product tank before being pumped to the regeneration area.

Major Processing Areas

The sodium/limestone double-alkali process is divided into the following process areas: (1) materials handling, (2) feed preparation, (3) gas handling, (4) SO_2 absorption, (5) reheat, (6) regeneration, (7) solids separation, and (8) fixation. Each of the processing areas is described in more detail below and an equipment list by area is shown in Table 13.

Materials Handling--

The materials-handling area consists of equipment to receive both limestone and soda ash by railcar or truck. The limestone is transferred to a 30-day storage pile with built-in live storage hoppers. The major limestone equipment consists of unloading conveyors, feed conveyors, bucket elevators, a dust collecting system, and feed bins. A scraper tractor is also included to manage the area. The unloading system has a capacity of 250 ton/hr and the feed system to the preparation area has a capacity of 100 ton/hr.

Dry soda ash is unloaded from trucks and stored as a slurry in a covered tank with 30-day storage capacity. The upper 10% of the storage tank contains a saturated Na_2CO_3 solution (32%-34% solids) over the slurry. This saturated solution is pumped to the thickener overflow tank where it is mixed with regenerated scrubbing solution to replace sodium losses in the system.

Feed Preparation--

The limestone feed preparation area consists of crushers, wet ball mills, mill product tanks with agitators and pumps, a slurry feed tank with an agitator and pumps, and a dust collection system. It is designed to produce a 60% solids slurry for addition to the regeneration area. The crushers and ball mills are situated in the limestone storage area, 1,500 feet from the FGD unit. The limestone is first crushed to 0 x 3/4 inch in two parallel 50% capacity gyratory crushers and then wet ground as a 60% solids slurry to 90% minus 325 mesh. Two operating ball mills with one spare mill are provided. The slurry from the wet ball mill is stored temporarily in the mill product tank (1-hour capacity) before being transferred to the limestone slurry feed tank which is located in the regeneration area. The feed tank has an 8-hour residence time.

The makeup soda ash solution is pumped directly from the storage tank and requires no further preparation equipment. It is added to the thickener overflow tank.

Gas Handling--

The gas-handling area consists of a feed plenum that distributes the flue gas to the individual absorber inlet ducts, the ductwork between the feed plenum and the stack plenum for each absorber train, two emergency bypass

TABLE 13. SODIUM/LIMESTONE DOUBLE-ALKALI PROCESS

EQUIPMENT LIST

Item (number): description	Total material cost, 1982 $
Area 1--Materials Handling	
1. Car shaker (1): Top mounting with crane, 20-hp shaker, 7-1/2-hp hoist	71,900
2. Car puller (1): 25-hp puller, 5-hp return	63,100
3. Hopper, limestone unloading (1): 16-ft diameter x 10-ft straight side, 2,570 ft3, carbon steel, 30-degree cone bottom	15,500
4. Feeder, limestone unloading (1): Vibrating pan, 30 in. wide x 60 in. long, 3-1/2 hp, 250 ton/hr	5,500
5. Conveyor, limestone unloading (1): Belt, 36 in. wide x 20 ft long, 5 hp, 250 ton/hr, 165 ft/min	11,400
6. Conveyor, limestone unloading (incline) (1): Belt, 36 in. wide x 310 ft long, 50 hp, 15-degree incline, 250 ton/hr, 165 ft/min	85,300
7. Dust collector, limestone unloading pit (1): Bag filter, polypropylene bag, 2,200 aft3/min, 7-1/2 hp, reverse jet cleaning	11,200
8. Pump, limestone pit sump (1): Duplex, 60 gpm, 70-ft head, 5 hp, carbon steel, neoprene lined	2,400
9. Conveyor, storage (1): Belt, 36 in. wide x 200 ft long, 5 hp, 250 ton/hr, 165 ft/min	73,100
10. Tripper, storage conveyor (1): 1 hp, 30 ft/min	27,200
11. Mobile equipment (1): Scraper tractor	141,900
12. Hopper, reclaim (2): 7-ft diameter x 4-1/4 ft deep x 2-ft bottom, 75 ft3, carbon steel, 60-degree cone bottom	2,400
13. Feeder, reclaim (2): Vibrating pan, 3-1/2 hp, 100 ton/hr	10,900

(Continued)

TABLE 13. (Continued)

Item (number): description	Total material cost, 1982 $
14. Dust collector, limestone reclaim pit (1): Bag filter, polypropylene bag, 2,200 aft3/min, 7-1/2 hp, reverse jet cleaning	7,800
15. Pump, reclaim pit sump (1): Duplex, 60 gpm, 70-ft head, 5 hp, carbon steel, neoprene lined	2,400
16. Conveyor, limestone reclaim (1): Belt, 30 in. wide x 200 ft long, 5 hp, 100 ton/hr, 105 ft/min	40,900
17. Conveyor, limestone reclaim (incline) (1): Belt, 30 in. wide x 193 ft long, 40 hp, 15-degree incline, 45-ft lift, 100 ton/hr, 105 ft/min	60,300
18. Elevator, live limestone feed (1): Continuous bucket, 14 in. x 8 in. x 11-3/4 in., 75 hp, 90-ft lift, 100 ton/hr	57,800
19. Conveyor, live limestone feed (1): Belt, 30 in. wide x 60 ft long, 7-1/2 hp, 100 ton/hr, 105 ft/min	20,500
20. Tripper, live limestone feed conveyor (1): 1 hp, 30 ft/min	27,200
21. Bin, crusher feed (3): 13-ft diameter x 21-ft straight side height, 3,100 ft3, w/cover, 50-degree cone bottom, carbon steel	43,300
22. Tank, soda ash storage (1): 36-ft diameter x 42 ft high, 275,000 gal, w/cover, carbon steel, insulated sparger in base	93,100
23. Pump, soda ash recirculating (2): Centrifugal, 200 gpm, 100-ft head, 7-1/2 hp, carbon steel, insulated (1 operating, 1 spare)	7,100
24. Pump, soda ash feed (2): Centrifugal, 30 gpm, 50-ft head, 1 hp, carbon steel, insulated (1 operating, 1 spare)	3,500
Total Area 1	885,700

(Continued)

TABLE 13. (Continued)

Item (number): description	Total material cost, 1982 $
Area 2--Feed Preparation	
1. Feeder, crusher (3): Weigh belt, 14 ft long, 2 hp	49,600
2. Crsher (3): Gyratory, 75 hp	297,100
3. Ball mill (3): Wet, open system, 9-ft diameter x 19 ft long, 553 hp, 233 ton/day (2 operating, 1 spare)	1,404,900
4. Dust collector, ball mill (3): Bag filter, polypropylene bag, 2,200 aft^3/min, 7-1/2 hp, reverse jet cleaning	23,300
5. Tank, mills product (3): 10-ft diameter x 10 ft high, 5,500 gal, open top, four 10-in. baffles, agitator supports, carbon steel, glass-filled polyester lining	13,700
6. Agitator, mills product tank (3): 40-in. diameter, 10 hp, neoprene coated	22,900
7. Pump, mills product tank (3): Centrifugal, 41 gpm, 60-ft head, 2 hp, carbon steel, neoprene lined (2 operating, 1 spare)	8,300
8. Tank, slurry feed (1): 19-1/2-ft diameter x 19-1/2 ft high, 43,100 gal, open top, four 20-in. baffles, agitator supports, carbon steel, glass-filled polyester lining	16,300
9. Agitator, slurry feed (1): 76-in. diameter, 40 hp, neoprene coated	34,900
10. Pump, slurry feed (8): Centrifugal, 20 gpm, 60-ft head, 1 hp, carbon steel, neoprene lined (4 operating, 4 spares)	21,500
Total Area 2	1,892,500

(Continued)

TABLE 13. (Continued)

Item (number): description	Total material cost, 1982 $
Area 3--Gas Handling	
1. Fans (5): Induced draft, 377,300 aft3/min, 10.0-in. static head, 850 hp, fluid drive, double width, double inlet, Inconel 625 (4 operating, 1 spare)	3,728,500
Total Area 3	3,728,500
Area 4--SO_2 Absorption	
1. SO_2 Absorber (5): Spray tower, 40 ft high x 34 ft wide x 18 ft deep, 1/4-in. carbon steel, neoprene lined, 316 stainless steel absorption trays, FRP chevron-vane entrainment separator (4 operating, 1 spare)	5,080,000
2. Recirculation tank (5): 15-ft diameter x 15 ft high, 19,100 gal, open top, four 15-in.-wide baffles, agitator supports, carbon steel, glass-filled polyester lining (4 operating, 1 spare)	47,400
3. Agitator, recirculation tank (5): 60-in. diameter, 8 hp, neoprene coated (4 operating, 1 spare)	54,300
4. Pump, presaturator (15): Centrifugal, 1,040 gpm, 100-ft head, 45 hp, carbon steel, neoprene lined (8 operating, 7 spares)	79,900
5. Pump, absorber bleed (10): Centrifugal, 1,660 gpm, 100-ft head, 75 hp, carbon steel, neoprene lined (4 operating, 6 spares)	120,600
6. Sootblowers (40): Air, fixed	111,900
Total Area 4	5,494,100

(Continued)

TABLE 13. (Continued)

Item (number): description	Total material cost, 1982 $
Area 5--Reheat	
1. Reheaters (5): Steam, tube type, 2,675 ft^2, six-tenths of tubes of Inconel 625 and four-tenths of Cor-Ten steel (4 operating, 1 spare)	2,607,800
2. Sootblowers (20): Air, retractable	183,000
Total Area 5	2,790,800
Area 6--Regeneration	
1. Reactor tank 1 (2): 25.8-ft diameter x 28.4 ft high, 110,800 gal, four 26-in. baffles, agitator supports, open top, carbon steel, glass-filled polyester lining	71,100
2. Reactor tank 2 (2): 26.6-ft diameter x 26.6 ft high, 110,800 gal, four 26-in. baffles, agitator supports, open top, carbon steel, glass-filled polyester lining	68,700
3. Reactor tank 3 (2): 27.6-ft diameter x 24.8 ft high, 110,800 gal, four 28-in. baffles, agitator supports, open top, carbon steel, glass-filled polyester lining	68,100
4. Reactor tank 4 (2): 28.7-ft diameter x 22.9 ft high, 110,800 gal, four 28-in. baffles, agitator supports, open top, carbon steel, glass-filled polyester lining	66,800
5. Agitator, reactor tank (8): 110-in. diameter, 21 hp, neoprene coated	298,000
6. Pump, reactor effluent (4): Centrifugal, 3,330 gpm, 60-ft head, 90 hp, carbon steel, neoprene lined (2 operating, 2 spares)	82,100
Total Area 6	654,800

(Continued)

TABLE 13. (Continued)

Item (number): description	Total material cost, 1982 $
Area 7--Solids Separation	
1. Thickener (1): 169-ft diameter x 9 ft high, carbon steel tank, concrete basin, includes 12-hp rake motor and mechanism	463,300
2. Tank, thickener overflow (1): 38-ft diameter x 9 ft high, 76,300 gal, open top, carbon steel, agitator supports, glass-filled polyester lining	26,900
3. Agitator, thickener overflow tank (1): 152-in. diameter, 6 hp, neoprene coated	16,600
4. Pump, scrubbing liquor return (10): Centrifugal, 1,735 gpm, 100-ft head, 75 hp, carbon steel, neoprene lined (4 operating, 6 spares)	108,600
5. Pump, thickener underflow (2): Centrifugal, 310 gpm, 60-ft head, 6 hp, carbon steel, neoprene lined (1 operating, 1 spare)	10,700
6. Tank, filter feed (1): 9.5-ft diameter x 9.5 ft high, 5,000 gal, open top, four 10-in. baffles, agitator supports, carbon steel, glass-filled polyester lining	4,500
7. Agitator filter feed tank (1): 40-in. diameter, 6 hp, neoprene coated	5,400
8. Pump, filter feed (3): Centrifugal, 155 gpm, 50-ft head, 4 hp, carbon steel, neoprene lined (2 operating, 1 spare)	11,400
9. Filter (3): Rotary vacuum, 8-ft diameter x 14-ft face, 85 total hp, includes auxiliary equipment (2 operating, 1 spare)	367,900
10. Pump, filtrate (2): Centrifugal, 422 gpm, 60-ft head, 10 hp, carbon steel, neoprene lined (1 operating, 1 spare)	10,900

(Continued)

TABLE 13. (Continued)

Item (number): description	Total material cost, 1982 $
11. Tank, filtrate surge (1): 9.2-ft diameter x 9.2 ft high, 4,600 gal, open top, carbon steel, neoprene lined	3,500
12. Pump, filtrate return (2): Centrifugal, 422 gpm, 60-ft head, 10 hp, carbon steel, neoprene lined (1 operating, 1 spare)	10,900
13. Conveyor, filter cake (1): Belt, 30 in. wide x 175 ft long, 1-1/2 hp, 100-ft incline, 50 ton/hr	37,100
Total Area 7	1,077,700
Area 8--Fixation	
1. Lime storage silo (1): 34-ft diameter x 34-ft straight side height, 30,500 ft^3, carbon steel, 60-degree cone bottom, includes feeder	54,600
2. Fly ash storage silo (1): 32-ft diameter x 32-ft straight side, 25,500 ft^3, carbon steel, 60-degree cone, includes feeder	49,300
3. Conveyor, fly ash and lime feed (1): Enclosed belt, 36 in. wide x 150 ft long, 1 hp, 50 ton/hr, 105 ft/min	35,300
4. Pugmill (2): Double agitator mixer, 20 hp, 304 stainless steel, 100 ton/hr	75,800
5. Conveyor, landfill product (1): Enclosed belt, 36 in. wide x 100 ft long, 1.5 hp, 100 ton/hr, 105 ft/min	28,100
6. Retaining wall (1): 8 ft high x 32 ft long, concrete	1,600

(Continued)

Table 13. (Continued)

Item (number): description	Total material cost, 1982 $
7. Dust collector (1): Bag filter, polypropylene bag, 2,200 aft3/min, 7-1/2 hp, reverse jet cleaning	7,800
Total Area 8	252,500

Basis: Most equipment cost estimates are based on informal vendor quotes and TVA information. These costs represent equipment costs only. Costs for piping, electrical equipment, instruments, foundations, and installation costs are not included. The differences in area costs between the equipment list and the capital summary sheets are due to these installation costs.

ducts around the absorber area (one from each end of the inlet plenum to each end of the stack plenum), and one Inconel 625 ID booster fan in each absorber train.

The ductwork upstream of the absorbers, the bypass ducts, and the ductwork downstream of the reheaters are constructed of Cor-Ten steel. The ductwork from the mist eliminator to and including the reheater casement is constructed of 316 stainless steel. Costs for the stack plenum are not included in the FGD system costs.

SO_2 Absorption--

The SO_2 absorption area contains five identical absorber trains, one of which is a spare. The absorbers are sieve tray towers constructed of neoprene-lined, 1/4-inch carbon steel. Each is equipped with a presaturator, two 316 stainless steel sieve trays, and a chevron-type mist eliminator. In the presaturator, scrubbing liquid at a L/G ratio of 2 gal/kaft^3 is sprayed into the incoming flue gas to cool and humidify the gas. Two banks of spray headers are included in the absorber for absorbent liquid distribution, one above the top tray spraying downward and one below the bottom tray spraying upward. This bottom spray provides a L/G of 1.0 gal/kaft^3 while the upper spray provides a L/G ratio of 5 gal/kaft^3. The design superficial gas velocity for the tower is 9 ft/sec.

The mist eliminator is mounted horizontally; it consists of three series of fiberglass, open-vane chevrons and reduces moisture entrainment to 0.1% (wt) of the flue gas. The mist eliminator is washed on an intermittent basis on the underside only.

Also included in the SO_2 absorption area is the absorbent liquid recirculation system consisting of tanks, agitators, piping, and centrifugal pumps. A flake-glass-lined recirculation tank (10-minute residence time) is situated directly beneath the absorber. The pumps are neoprene-lined carbon steel.

Reheat--

Each absorber train contains an indirect steam reheater between the mist eliminator and the booster fan. They increase flue gas temperature from 125°F to 170°F, which with a 5°F temperature increase through the booster fan provides a 175°F gas temperature at the entrance to the stack. About 60% of the tubes on the upstream side, where the gas temperature is below 150°F, are made of Inconel 625. The rest are Cor-Ten steel. Retractable sootblowers are included in the reheater area to clean the tube banks. Steam is supplied to the reheater at 500 psig and 470°F.

Regeneration--

The regeneration area has two parallel trains of four reactor tanks in series. Each tank has a 15-minute residence time. The absorber bleedstream and limestone slurry are fed to the first tank and overflows by gravity to each of the succeeding tanks. The tanks are carbon steel with a glass-filled polyester lining and are equipped with agitators. The effluent from the final reactor tank is pumped to the thickener.

Solids Separation--

This area includes a single gravity thickener which concentrates the solids from 1.4% to 25% and three rotary drum vacuum filters (two operating and one spare) which produce a 55% solids cake. Also included in the area are tanks, pumps, and agitators to provide surge capacity around the filters and a conveyor to transport the filter cake to the pugmills in the fixation area.

Fixation--

The fixation area includes two 304 stainless steel pugmills (one operating, one spare); two carbon steel storage silos, one for lime (15-day supply) and the other for fly ash (2-day supply); a conveyor for transporting the lime and fly ash to the pugmills; and an enclosed conveyor for transferring the fixed waste to the loading pit. The equipment in this area is designed to handle up to 75 tons of fixed waste per hour. The fixed waste (with a dry bulk density of 85 lb/ft^3) is trucked to the onsite landfill.

FORCED-OXIDATION LIMESTONE PROCESS

The forced-oxidation limestone process included for comparison purposes is based in part on testing at the Shawnee test facility at TVA's Shawnee Steam Plant (30) and has been used in other TVA economic evaluations (29). A flowsheet and a material balance are shown in Figures 3 and 4 and Table 14.

Process Description and Principles of Operation

The FGD plenum distributes flue gas from the boiler ID fans to the four operating spray tower absorber trains. Each absorber has a presaturator and contains four spray banks and three gas distribution grids. The flue gas is cooled to 125°F with a spray of limestone slurry in the presaturator as it enters the absorber. The flue gas from the absorber passes through a mist eliminator. The saturated flue gas is reheated to 170°F by an inline steam reheater and passes through an ID booster fan, which further heats the gas to 175°F by compression before it enters the stack.

Slurry from the absorber flows into an oxidation tank directly below the absorber. Compressed air is sparged into the slurry to oxidize the calcium sulfite formed in the absorber to calcium sulfate.

The recirculation tank is fed by an overflow from the oxidation tank. Fresh limestone slurry and process water are added to control the undissolved solids level at 8%. Slurry from the recirculating tank is pumped to both the presaturator and the absorber.

A bleedstream is pumped from the oxidation tank to the solids separation area where the slurry is concentrated from 8% to 40% solids in a gravity thickener and then to an 85% solids filter cake by rotary vacuum drum filters. The filter cake is trucked to an onsite landfill for disposal.

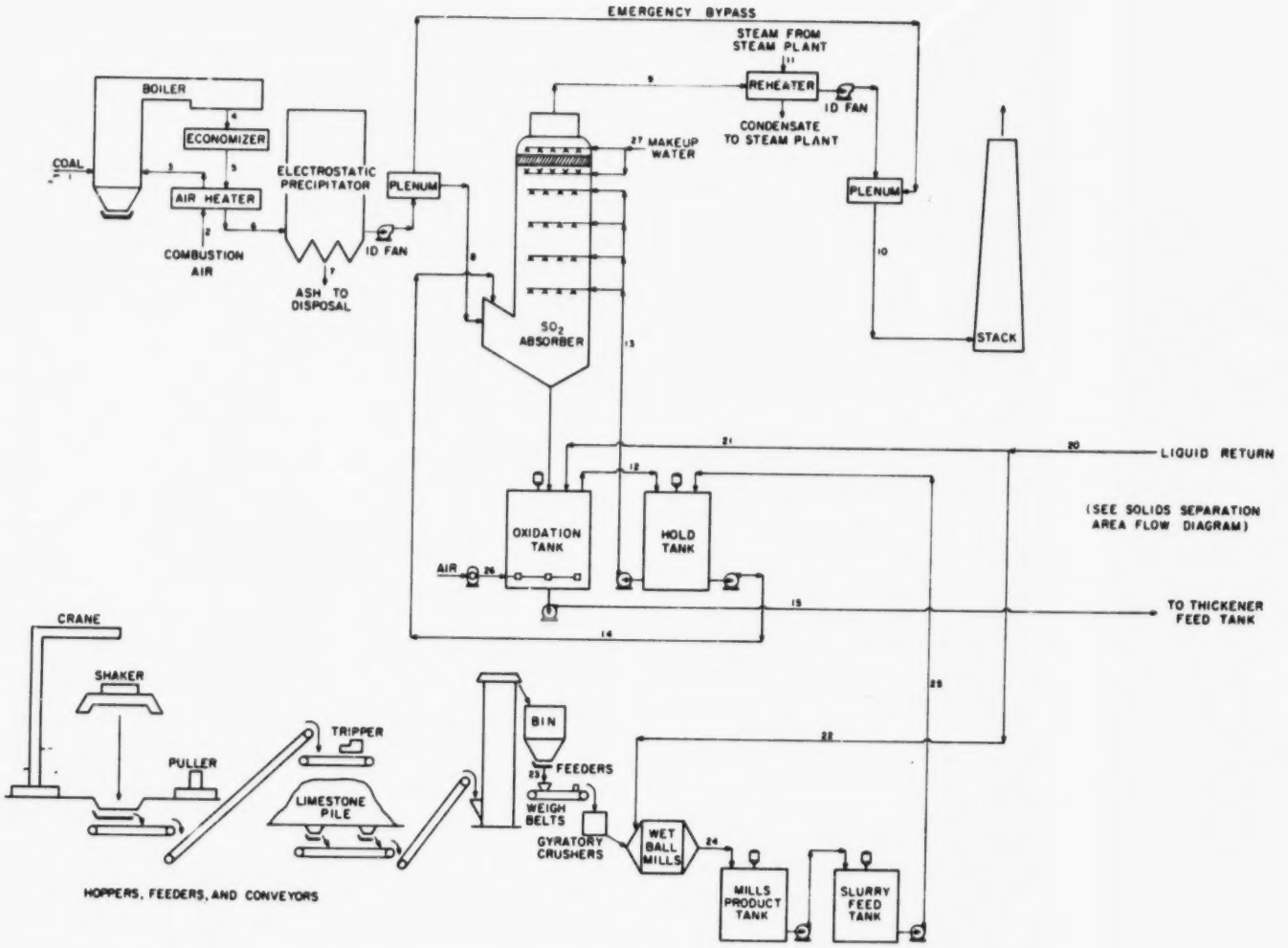

Figure 3. Forced-oxidation limestone process flow diagram.

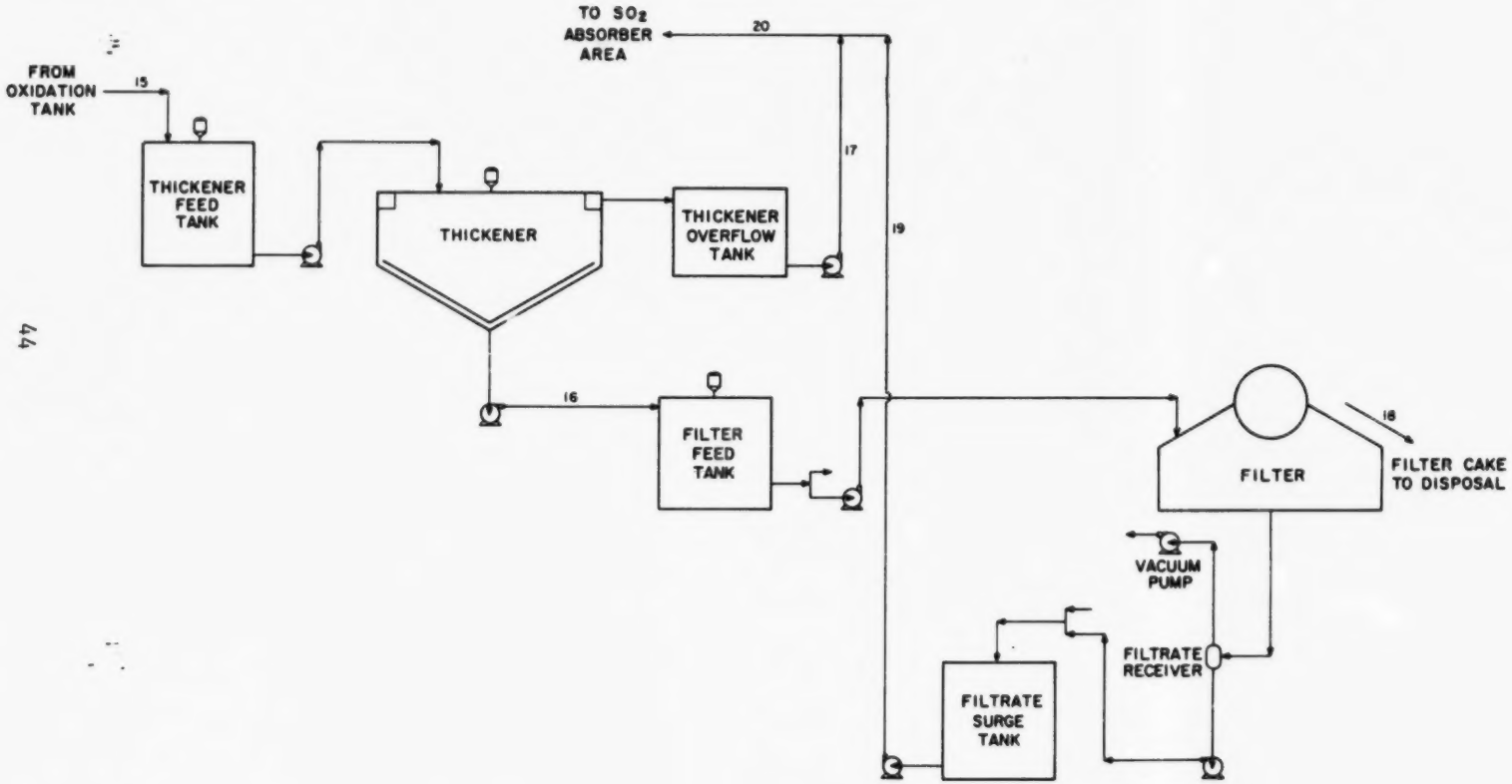

Figure 4. Forced-oxidation limestone process solids separation area flow diagram.

TABLE 14. FORCED-OXIDATION LIMESTONE PROCESS

MATERIAL BALANCE

Stream No.		1	2	3	4	5
	Description	Coal to boiler	Total air to air heater	Combustion air to boiler	Gas to economizer	Gas to air heater
1	Total stream, lb/hr	406,000	5,072,000	4,378,000	4,772,000	4,772,000
2						
3	Flow rate, ft^3/min@60°F		1,120,000	967,300	1,003,000	1,003,000
4	Temperature, °F		80	495	705	705
5	Pressure, psig					
6	Flow rate, gal/min					
7	Specific gravity					
8	pH					
9	Undissolved solids, %					
10						

Stream No.		6	7	8	9	10
	Description	Gas to electrostatic precipitator	Fly ash to disposal	Gas to presaturator	Gas to reheater	Gas to stack
1	Total stream, lb/hr	5,465,000	50,050	5,415,000	5,639,000	5,639,000
2						
3	Flow rate, ft^3/min@60°F	1,157,000		1,156,000	1,234,000	1,234,000
4	Temperature, °F	300		300	124	175
5	Pressure, psig					
6	Flow rate, gal/min					
7	Specific gravity					
8	pH					
9	Undissolved solids, %					
10						

Stream No.		11	12	13	14	15
	Description	Steam to reheater	Oxidation tank overflow	Recycle slurry to absorber	Recycle slurry to presaturator	Absorber slurry to thickener
1	Total stream, lb/hr	98,580	80,100,000	77,270,000	2,916,000	948,900
2						
3	Flow rate, ft^3/min@60°F					
4	Temperature, °F	470				
5	Pressure, psig	500				
6	Flow rate, gal/min		152,600	147,000	5,547	1,805
7	Specific gravity		1.05	1.05	1.05	1.05
8	pH					
9	Undissolved solids, %		8	8	8	8
10						

Stream No.		16	17	18	19	20
	Description	Thickener bottoms to filter	Thickener overflow return	Gypsum cake to disposal	Filtrate return	Clear liquid return
1	Total stream, lb/hr	189,800	731,600	89,300	100,900	832,100
2						
3	Flow rate, ft^3/min@60°F					
4	Temperature, °F					
5	Pressure, psig					
6	Flow rate, gal/min	286	1,464		201	1,665
7	Specific gravity	1.3	1.0	2.0	1.0	1.0
8	pH					
9	Undissolved solids, %	40		85		
10						

(Continued)

TABLE 14. (Continued)

	Stream No.	21	22	23	24	25
	Description	Clear liquid to oxidation tank	Clear liquid to wet ball mill	Limestone to weigh feeder	Slurry to mills product tank	Limestone slurry to recirculation tank
1	Total stream, lb/hr	800,000	32,130	51,950	86,580	86,580
2						
3	Flow rate, ft^3/min@60°F					
4	Temperature, °F					
5	Pressure, psig					
6	Flow rate, gal/min	1,601	64		109	109
7	Specific gravity	1.0	1.0		1.6	1.6
8	pH					
9	Undissolved solids, %				60	60
10						

	Stream No.	26	27			
	Description	Air to oxidizer	Makeup water to absorber			
1	Total stream, lb/hr	59,950	286,900			
2						
3	Flow rate, ft^3/min@60°F	13,060				
4	Temperature, °F					
5	Pressure, psig					
6	Flow rate, gal/min		574			
7	Specific gravity		1.0			
8	pH					
9	Undissolved solids, %					
10						

1						
2						
3						
4						
5						
6						
7						
8						
9						
10						

1						
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9						
10						

The thickener overflow and the filtrate are returned to the process. A portion of the process water is pumped to the ball mill for preparation of fresh limestone slurry. The rest is returned to the recirculation tank in the absorption area.

Major Processing Areas

The limestone-scrubbing process is divided into the following areas: (1) materials handling, (2) feed preparation, (3) gas handling, (4) SO_2 absorption, (5) oxidation, (6) reheat, and (7) solids separation. Each of these processing areas is described in more detail below and an equipment list is shown in Table 15.

Materials Handling--

The materials-handling area consists of equipment to receive and unload 0- x 1-1/2-inch limestone from a railcar or truck and transfer it to a 30-day storage pile with built-in live storage hoppers. The major equipment consists of unloading conveyors, feed conveyors, bucket elevators, a dust collecting system, and feed bins. A scraper tractor is also included to manage the area. The unloading system has a capacity of up to 250 ton/hr and the feed system to the preparation area has a capacity of up to 100 ton/hr.

Feed Preparation--

The feed preparation area for the limestone system consists of crushers, wet ball mills, mill product tanks with agitators and pumps, a slurry feed tank with an agitator and pumps, and a dust collection system. The feed preparation unit is designed to produce a 60% solids slurry. The crushers and ball mills are situated in the limestone storage area, 1,500 feet from the FGD unit. The limestone is first crushed to 0 x 3/4 inch in two parallel 50% capacity gyratory crushers and then wet ground as a 60% solids slurry to 90% minus 325 mesh. Two operating mills and one spare are provided. The slurry from the wet ball mill is temporarily stored in the mills product tank. The slurry is then pumped to the slurry feed tank which is located at the FGD unit. This tank has an 8-hour capacity and is equipped with pumps for feeding the slurry to the absorber hold tanks.

Gas Handling--

The gas-handling area consists of a feed plenum that distributes the flue gas to the individual absorber inlet ducts, the absorber ductwork between the feed plenum and the stack plenum, two emergency bypass ducts around the absorber area (one from the inlet plenum to each end of the stack plenum), and one Inconel 625 ID booster fan per absorber train. The ductwork upstream of the absorbers, the bypass ducts, and the ductwork downstream of the reheaters are constructed of Cor-Ten steel. The ductwork from the mist eliminator up to and including the reheater casement is constructed of 316 stainless steel. Costs for the stack plenum are not included in the FGD system costs.

SO_2 Absorption--

The SO_2 absorption area contains five identical trains of processing equipment, four of which are operating and one is a spare. The spray tower

TABLE 15. FORCED-OXIDATION LIMESTONE PROCESS

EQUIPMENT LIST

Item (number): description	Total material cost, 1982 $
Area 1--Materials Handling	
1. Car shaker (1): Top mounting with crane, 20-hp shaker, 7-1/2-hp hoist	71,900
2. Car puller (1): 25-hp puller, 5-hp return	63,100
3. Hopper, limestone unloading (1): 16-ft diameter x 10-ft straight side, 2,570 ft3, carbon steel, 30-degree cone bottom	15,500
4. Feeder, limestone unloading (1): Vibrating pan, 30 in. wide x 60 in. long, 3-1/2 hp, 250 ton/hr	5,500
5. Conveyor, limestone unloading (1): Belt, 36 in. wide x 20 ft long, 5 hp, 250 ton/hr, 165 ft/min	11,400
6. Conveyor, limestone unloading (incline) (1): Belt, 36 in. wide x 310 ft long, 50 hp, 15-degree incline, 250 ton/hr, 165 ft/min	85,300
7. Dust collector, limestone unloading pit (1): Bag filter, polypropylene bag, 2,200 aft3/min, 7-1/2 hp, reverse jet cleaning	11,200
8. Pump, limestone pit sump (1): Duplex, 60 gpm, 70-ft head, 5 hp, carbon steel, neoprene lined	2,400
9. Conveyor, storage (1): Belt, 36 in. wide x 200 ft long, 5 hp, 250 ton/hr, 165 ft/min	73,100
10. Tripper, storage conveyor (1): 1 hp, 30 ft/min	27,200
11. Mobile equipment (1): Scraper tractor	141,900
12. Hopper, reclaim (2): 7-ft diameter x 4-1/4 ft deep x 2-ft bottom, 75 ft3, carbon steel, 60-degree cone bottom	2,400

(Continued)

TABLE 15. (Continued)

Item (number): description	Total material cost, 1982 $
13. Feeder, reclaim (2): Vibrating pan, 3-1/2 hp, 100 ton/hr	10,900
14. Dust collector, limestone reclaim pit (1): Bag filter, polypropylene bag, 2,200 aft^3/min, 7-1/2 hp, reverse jet cleaning	7,800
15. Pump, reclaim pit sump (1): Duplex, 60 gpm, 70-ft head, 5 hp, carbon steel, neoprene lined	2,400
16. Conveyor, limestone reclaim (1): Belt, 30 in. wide x 200 ft long, 5 hp, 100 ton/hr, 105 ft/min	40,900
17. Conveyor, limestone reclaim (incline) (1): Belt, 30 in. wide x 200 ft long, 40 hp, 15-degree incline, 45-ft lift, 100 ton/hr, 100 ft/min	60,300
18. Elevator, live limestone feed (1): Continuous bucket, 14 in. x 8 in. x 11-3/4 in., 75 hp, 90-ft lift, 100 ton/hr	57,800
19. Conveyor, live limestone feed (1): Belt, 30 in. wide x 60 ft long, 7-1/2 hp, 100 ton/hr, 100 ft/min	20,500
20. Tripper, live limestone feed conveyor (1): 1 hp, 30 ft/min	27,200
21. Bin, crusher feed (3): 13-ft diameter x 21-ft straight side height, 3,100 ft^3, w/cover, 50-degree cone bottom, carbon steel (2 operating, 1 spare)	43,300
Total Area 1	782,000
Area 2--Feed Preparation	
1. Feeder, crusher (3): Weigh belt, 14 ft long, 2 hp (2 operating, 1 spare)	49,600

(Continued)

TABLE 15. (Continued)

Item (number): description	Total material cost, 1982 $
2. Crusher (3): Gyratory, 75 hp (2 operating, 1 spare)	297,100
3. Ball mill (3): Wet, open system, 9-1/2-ft diameter x 19 ft long, 740 hp, 325 ton/day (2 operating, 1 spare)	1,721,200
4. Dust collector, ball mill (3): Bag filter, polypropylene bag, 2,200 aft^3/min, 7-1/2 hp, reverse jet cleaning (2 operating, 1 spare)	23,300
5. Tank, mills product (3): 10-ft diameter x 10 ft high, 5,500 gal, open top, four 10-in. baffles, agitator supports, carbon steel, glass-filled polyester lining (2 operating, 1 spare)	13,700
6. Agitator, mills product tank (3): 40-in. diameter, 10 hp, neoprene coated (2 operating, 1 spare)	22,900
7. Pump, mills product tank (3): Centrifugal, 55 gpm, 60-ft head, 2 hp, carbon steel, neoprene lined (2 operating, 1 spare)	8,500
8. Tank, slurry feed (1): 21-ft diameter x 21 ft high, 57,700 gal, open top, four 21-in. baffles, agitator supports, carbon steel, glass-filled polyester lining	19,800
9. Agitator, slurry feed (1): 84-in. diameter, 49 hp, neoprene coated	43,600
10. Pump, slurry feed (8): Centrifugal, 27 gpm, 60-ft head, 1 hp, carbon steel, neoprene lined (4 operating, 4 spares)	21,500
Total Area 2	2,221,200

(Continued)

TABLE 15. (Continued)

Item (number): description	Total material cost, 1982 $
Area 3--Gas Handling	
1. Fans (5): Induced draft, 377,300 aft3/min, 7.9-in. static head, 663 hp, fluid drive, double width, double inlet, Inconel 625 (4 operating, 1 spare)	3,401,300
Total Area 3	3,401,300
Area 4--SO_2 Absorption	
1. SO_2 Absorber (5): Spray tower, 40 ft high x 34 ft wide x 17 ft deep, 1/4-in. carbon steel, neoprene lined, 316 stainless steel grids, FRP, chevron-vane entrainment separator (4 operating, 1 spare)	5,517,200
2. Recirculation tank (5): 38-ft diameter x 38 ft high, 323,400 gal, open top, four 38-in.-wide baffles, agitator supports, carbon steel, glass-filled polyester lining (4 operating, 1 spare)	358,500
3. Agitator, recirculation tank (5): 152-in. diameter, 66 hp, neoprene coated (4 operating, 1 spare)	442,800
4. Pump, presaturator (10): Centrifugal, 1,400 gpm, 100-ft head, 61 hp, carbon steel, neoprene lined (4 operating, 6 spares)	103,200
5. Pump, recirculation (15): Centrifugal, 18,400 gpm, 100-ft head, 813 hp, carbon steel, neoprene lined (8 operating, 7 spares)	1,689,800
6. Pump, makeup water (2): Centrifugal, 3,470 gpm, 200-ft head, 292 hp, carbon steel (1 operating, 1 spare)	33,200
7. Sootblowers (40): Air, fixed	111,900
Total Area 4	8,256,600

(Continued)

TABLE 15. (Continued)

Item (number): description	Total material cost, 1982 $
Area 5--Reheat	
1. Reheaters (5): Steam, tube type, 2,750 ft2, 60% of tubes made of Inconel 625 and 40% made of Cor-Ten steel (4 operating, 1 spare)	2,638,800
2. Sootblowers (20): Air, retractable	183,000
Total Area 5	2,821,800
Area 6--Oxidation	
1. Tank, oxidation (5): 30-ft diameter x 38 ft high, 202,100 gal, open top, four 30-in.-wide baffles, agitator supports, carbon steel, glass-filled polyester lining, includes air sparger (4 operating, 1 spare)	366,700
2. Agitator, oxidation tank (5): 120-in. diameter, 59 hp, neoprene coated (4 operating, 1 spare)	317,600
3. Pump, oxidation tank (8): Centrifugal, 450 gpm, 60-ft head, 12 hp, carbon steel, neoprene lined (4 operating, 4 spares)	37,700
4. Blower, oxidation air (6): 3,264 sft3/min, 322 hp (4 operating, 2 spares)	208,400
Total Area 6	930,400
Area 7--Solids Separation and Disposal	
1. Tank, thickener feed (1): 19-ft diameter x 38 ft high, 85,760 gal, open top, four 19-in.-wide baffles, agitator supports, carbon steel, glass-filled polyester lining	28,500

(Continued)

TABLE 15. (Continued)

Item (number): description	Total material cost, 1982 $
2. Agitator, thickener feed tank (1): 80-in. diameter, 41 hp, neoprene coated	31,300
3. Pump, thickener feed (2): Centrifugal, 1,800 gpm, 60-ft head, 48 hp, carbon steel, neoprene lined (1 operating, 1 spare)	19,200
4. Thickener (1): 48-ft diameter x 5-1/2 ft high, carbon steel sides, concrete basin, includes 1-hp rake motor and mechanism	85,500
5. Tank, thickener overflow (1): 27-1/2-ft diameter x 5-1/2 ft high, 24,150 gal, open top, carbon steel	6,600
6. Pump, thickener overflow tank (2): Centrifugal, 1,464 gpm, 75-ft head, 46 hp, carbon steel, neoprene lined (1 operating, 1 spare)	12,900
7. Pump, thickener underflow (2): Centrifugal, 286 gpm, 9-ft head, 1 hp, carbon steel, neoprene lined (1 operating, 1 spare)	7,800
8. Tank, filter feed (1): 9-ft diameter x 9 ft high, 4,725 gal, open top, four 9-in.-wide baffles, agitator supports, carbon steel, glass-filled polyester lining	3,700
9. Agitator, filter feed tank (1): 36-in. diameter, 7 hp, neoprene coated	5,600
10. Pump, filter feed tank (3): Centrifugal, 143 gpm, 50-ft head, 4 hp, carbon steel, neoprene lined (2 operating, 1 spare)	11,900
11. Filter (3): Rotary vacuum, 8-ft diameter x 14-ft face, 85 total hp, includes auxiliary equipment (2 operating, 1 spare)	380,800
12. Pump, filtrate (4): Centrifugal, 100 gpm, 20-ft head, 1 hp, carbon steel, neoprene lined (2 operating, 2 spares)	17,300

(Continued)

TABLE 15. (Continued)

Item (number): description	Total material cost, 1982 $
13. Tank, filtrate surge (1): 8-ft diameter x 8 ft high, 3,320 gal, open top, carbon steel	1,700
14. Pump, filtrate surge tank (2): Centrifugal, 200 gpm, 85-ft head, 7 hp, carbon steel, neoprene lined (1 operating, 1 spare)	9,300
15. Conveyor, filter cake (incline) (1): Belt, 30 in. wide x 175 ft long, 1-1/2 hp, 100-ft incline, 50 ton/hr, 105 ft/min	37,100
16. Retaining wall, gypsum (1): 8 ft high x 32 ft long, concrete	1,600
Total Area 7	660,800

Basis: Most equipment cost estimates are based on informal vendor quotes and TVA information. These costs represent equipment costs only. Costs for piping, electrical equipment, instruments, foundations, and other installation costs are not included. The differences in area costs between the equipment list and the capital summary sheets are due to these installation costs.

absorbers are constructed of neoprene-lined, 1/4-inch carbon steel. Each is equipped with a presaturator to cool and humidify the flue gas by spraying slurry at a L/G ratio of 4 gal/kaft^3. Each also includes a chevron-type mist eliminator. Three 316L stainless steel grids are provided in the spray tower to control gas distribution. Four banks of sprays are used for liquid distribution, one above each of the top three grids spraying downward and one below the bottom grid spraying upward. The design L/G ratio is 106 gal/kaft^3 and the superficial gas velocity is 10 ft/sec.

A horizontal, three-pass, open-vane, fiberglass, chevron-type mist eliminator above the absorbers is provided to reduce the entrained moisture content of the scrubbed gas to a maximum level of 0.1% (by wt) of the flue gas. The mist eliminator is continuously washed on the underside and intermittently on the topside with fresh makeup water.

Also included in the SO_2 absorption area is the absorbent liquid recirculation system (excluding the oxidation tank and equipment) consisting of tanks, agitators, piping, and pumps. The recirculation tank has a 10-minute residence time. All hold tanks are baffled, agitated, and made of carbon steel with glass-filled polyester linings. The pumps are neoprene-lined carbon steel centrifugal pumps.

Oxidation--

Major equipment in the oxidation area consists of low-pressure air compressors to supply air to the oxidation tank, the oxidation tank itself, a sparging ring in the base of the tank, an agitator to maintain solids dispersion and improve air distribution, and a bleed pump to transfer the oxidized slurry to the waste disposal area. The oxidation tank has a 5-minute residence time and is situated directly under the SO_2 absorber.

Reheat--

Each absorber train has an inline steam reheater which raises the flue gas temperature from 125°F to 170°F. (The remaining 5°F of reheat is obtained by compression in the ID booster fans.) About six-tenths of the tubes on the upstream side where the flue gas temperature is below 150°F is constructed of Inconel 625. The remaining tubes are Cor-Ten steel. The reheater is equipped with retractable sootblowers. Steam is supplied to the reheater at 500 psig and 470°F.

Solids Separation--

The solids separation area includes a gravity thickener and three vacuum filters (two operating and one spare) which produce an 85% (wt) solids filter cake. Other equipment consists of a thickener feed tank and agitator, pumps, thickener overflow tank and pumps, variable-speed thickener underflow pumps, a filter feed tank with an agitator and pumps, filter cake conveyor, and filtrate surge tank and pumps.

RESULTS

Based on the design and economic premises and the process designs as described in the previous sections, capital investments in mid-1982 dollars and first-year annual revenue requirements (which include levelized capital charges) and levelized annual revenue requirements in mid-1984 dollars were prepared for the sodium/limestone double-alkali and the forced-oxidation limestone processes. The accuracy associated with this type of preliminary-grade cost estimate is -15%, +30%. It is the potential difference in costs for an actual installation of similar design in comparison with the estimated cost, expressed as a percent of the estimated cost. When comparing the preliminary-grade costs for processes evaluated using the same procedures, the range of significant difference is about ±10%.

CAPITAL INVESTMENT

The capital investments for the sodium/limestone double-alkali and forced-oxidation limestone processes are shown in Table 16. The total capital investment for the double-alkali process is $95 million (190 $/kW) as compared with $103 million (206 $/kW) for the limestone process. Direct investment, representing installed equipment costs, is $47 million for the double-alkali process and $57 million for the limestone process. In some process areas, the costs of the two processes are similar. In the raw materials-handling and feed preparation areas, 40% less limestone is processed for the double-alkali process because of its lower stoichiometry (1.0 versus 1.4 for the limestone process) but the soda ash storage and handling equipment essentially offsets this advantage, making the total costs for these areas similar for both processes. The reheat area costs are also similar for both processes.

The largest cost difference is in the SO_2 absorption area where the double-alkali process is $9 million less expensive. The absorbers of both processes are similar in size (the double-alkali process absorbers are actually slightly larger in the cross-sectional area because of the lower gas velocity used) but the greater complexity of the limestone process scrubbing-liquid system makes the limestone process absorption area slightly more expensive. Most of the lower SO_2 absorption costs for the double-alkali process are a result of its smaller hold tank and, particularly, its low L/G ratio. The double-alkali process has 19,000-gallon recirculation tanks with 8-hp agitators because no reactions occur in them and there is no slurry requiring agitation to keep it in suspension. The limestone process has 323,000-gallon recirculation tanks to allow time for limestone dissolution and reaction and

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TABLE 16. CAPITAL INVESTMENT

	Capital investment, k$	
	Double alkali	Limestone
Direct Investment		
Materials handling	2,426	2,528
Feed preparation	4,506	4,715
Gas handling	10,800	11,281
SO_2 absorption	11,348	20,288
Reheat	3,630	3,634
Regeneration	1,506	0
Oxidation	0	2,670
Solids separation	5,493	3,679
Fixation	906	0
Total process capital	40,615	48,795
Services, utilities, and miscellaneous	2,437	2,928
Total direct investment excluding landfill	43,052	51,723
Landfill construction	5,247	3,781
Landfill equipment	1,454	1,123
Landfill credit (fly ash disposal)	(2,312)	0
Total direct investment	47,441	56,627
Indirect Investment		
Engineering design and supervision	3,014	3,621
Architect and engineering contractor	861	1,034
Construction expense	6,888	8,276
Contractor fees	2,153	2,586
Contingency	11,194	6,724
Waste disposal indirect investment	1,720	1,707
Total fixed investment	73,271	80,575
Other Capital Investment		
Allowance for startup and modification	6,716	5,917
Interest during construction	11,430	12,570
Royalties	406	0
Land	554	614
Working capital	2,775	3,392
Total capital investment	95,152	103,068
Dollars per kW of generation capacity	190.30	206.14

Basis: 500-MW new coal-fired power unit; 3.5% in coal; 89% SO_2 removal; fixation and landfill solids disposal; north-central, new power unit with 30-yr life at 5,500 hr/yr full-load operation; 1 spare absorber train, and provisions for emergency bypass of 50% of the scrubbed gas; reheat to 175°F as required.

66-hp agitators to keep the slurry in suspension. The double-alkali process also has a very low liquid recirculation rate from the recirculation tank; a L/G ratio of 3 is required for the presaturator and bottom tray (the remaining scrubbing liquid is pumped from the regeneration area). The limestone process has a L/G ratio of 106 and has recirculation pump costs of $1.7 million, as compared with $80,000 for the double-alkali process. The double-alkali process also has slightly lower ductwork costs because the smaller recirculation tanks permit a more compact design.

The complicated 2-train, multiple-tank regeneration area of the double-alkali process, necessary for the use of limestone, does not add materially to the process costs. The regeneration area direct investment is $1.5 million, less than 4% of the total process direct investment. In comparison, the regeneration area cost for a sodium/lime process that was evaluated in another TVA study (5) was 1.3% of the total process direct investment.

The solids separation (i.e., dewatering) costs for the double-alkali process are about 50% higher than those of the limestone process because of the high-sulfite waste it produces. Including the costs of forced oxidation--which produces the lower solids separation costs--in the limestone process, however, the total costs of preparing the waste for disposal are $6.4 million for both the double-alkali process (solids separation and fixation) and the limestone process (oxidation and solids separation). Including the credit to the double-alkali process for fly ash disposal, the landfill disposal costs are about 10% less for the double-alkali process. This is partly due to the absence of unreacted limestone in the double-alkali process waste and partly due to the lower unit costs in terms of waste quantities for larger volume landfills.

Indirect investment and other capital investment, which are based for the most part on direct capital investment, account for the remaining difference in capital investment between the two processes. Exceptions to the generally lower indirect investment and other capital investment costs for the double-alkali process are the contingency fee and the allowance for startup and modifications, which are higher because of the lower level of technical development of the double-alkali process.

Overall, the lower capital investment of the double-alkali process is a result of the use of a highly efficient scrubbing solution, which greatly reduces the cost of the liquid-handling portion of the SO_2 absorption system--but not of the absorbers themselves. The additional costs for storage and handling of the soda ash are minimal. The additional costs for the larger and more complicated regeneration area necessary for the use of limestone have little effect on the overall costs.

ANNUAL REVENUE REQUIREMENTS

The annual revenue requirements for the double-alkali process are shown in Table 17 and for the limestone process in Table 18. The first-year annual

TABLE 17. SODIUM/LIMESTONE DOUBLE-ALKALI PROCESS

ANNUAL REVENUE REQUIREMENTS

	Annual quantity	Unit cost, $	Total annual cost, k$
Direct Costs - First Year			
Raw materials			
Limestone	107,700 tons	8.50/ton	915
Soda ash	6,190 tons	160.00/ton	990
Lime	10,580 tons	75.00/ton	794
Total raw materials cost			2,699
Conversion costs			
Operating labor and supervision			
FGD	43,860 man-hr	15.00/man-hr	658
Solids disposal	33,280 man-hr	21.00/man-hr	699
Utilities			
Steam	528,500 klb	2.50/klb	1,321
Process water	262,100 kgal	0.14/kgal	37
Electricity	29,618,000 kWh	0.037/kWh	1,096
Diesel fuel	123,200 gal	1.60/gal	197
Landfill fly ash credit			(312)
Maintenance			
Labor and material			2,715
Analysis	4,990 man-hr	21.00/man-hr	105
Total conversion costs			6,516
Total direct costs			9,215
Indirect Costs - First Year			
Overheads			
Plant and administrative (60% of conversion costs less utilities)			2,319
Total first-year operating and maintenance costs			11,534
Levelized capital charges (14.7% of total capital investment)			13,987
Total first-year annual revenue requirements			25,521
Levelized first-year operating and maintenance costs (1.886 first-year O and M)			21,753
Levelized capital charges (14.7% of total capital investment)			13,987
Levelized annual revenue requirements			35,740

	M$	Mills/kWh
First-year annual revenue requirements	25.52	9.28
Levelized annual revenue requirements	35.74	13.00

Basis: One-year, 5,500-hr operation of system as described in capital investment table. Mid-1984 cost basis.

TABLE 18. FORCED-OXIDATION LIMESTONE PROCESS

ANNUAL REVENUE REQUIREMENTS

	Annual quantity	Unit cost, $	Total annual cost, k$
Direct Costs - First Year			
Raw materials			
Limestone	142,900 tons	8.50/ton	1,214
Total raw materials cost			1,214
Conversion costs			
Operating labor and supervision			
FGD	43,860 man-hr	15.00/man-hr	658
Waste disposal	29,120 man-hr	21.00/man-hr	611
Utilities			
Steam	542,200 klb	2.50/klb	1,356
Process water	193,400 kgal	0.14/kgal	27
Electricity	57,657,000 kWh	0.037/kWh	2,133
Diesel fuel	103,200 gal	1.60/gal	165
Maintenance			
Labor and material			4,285
Analysis	4,990 man-hr	21.00/man-hr	105
Total conversion costs			9,340
Total direct costs			10,554
Indirect Costs - First Year			
Overheads			
Plant and administrative (60% of conversion costs less utilities)			3,395
Total first-year operating and maintenance costs			13,949
Levelized capital charges (14.7% of total capital investment)			15,151
Total first-year annual revenue requirements			29,100
Levelized first-year operating and maintenance costs (1.886 first-year O and M)			26,309
Levelized capital charges (14.7% of total capital investment)			15,151
Levelized annual revenue requirements			41,460

	M$	Mills/kWh
First-year annual revenue requirements	29.10	10.58
Levelized annual revenue requirements	41.46	15.08

Basis: One-year, 5,500-hr operation of system as described in capital investment table. Mid-1984 cost basis.

revenue requirements for the double-alkali process are $26 million (9.28 mills/kWh), as compared with $29 million (10.58 mills/kWh) for the limestone process. The annual direct costs for the double-alkali process are $9.2 million and constitute 36% of the total first-year annual revenue requirements. The annual direct costs for the limestone process are $10.6 million and also constitute 36% of the total annual revenue requirements. The remaining costs are overheads based on the direct costs and capital charges based on the capital investment.

The levelized annual revenue requirements are $35.7 million (13.00 mills/kWh) for the double-alkali process and $41 million (15.08 mills/kWh) for the limestone process. These take into account a projected estimate of the value of money and inflation rate as explained in the premises. They are sometimes regarded as a more accurate representation of the actual operating costs over the life of the project than first-year or average-year costs.

A comparison of the major components of the first-year annual revenue requirements is shown in Table 19. Among the direct costs, the major costs for the double-alkali process, in order of importance, are maintenance, raw materials, labor, steam, and electricity. For the limestone-scrubbing process, they are maintenance, electricity, steam, labor, and raw materials. The major differences in direct costs between the processes are in maintenance, electricity, and raw material costs. Maintenance costs for the sodium/limestone double-alkali process are 37% lower than for the forced-oxidation limestone process as a result of the forced-oxidation limestone process having high maintenance costs associated with circulating large volumes of slurry. This also results in electricity costs for the sodium/limestone double-alkali process being about half those of the forced-oxidation limestone process. Raw material costs are over twice as high for the double-alkali process, however, and are divided about equally between costs for limestone, soda ash, and lime for fixation.

Overall, the double-alkali process direct costs are about 13% lower than those of the limestone process because of lower maintenance and electricity costs, both a result of using a highly reactive scrubbing solution instead of a limestone slurry. With the lower overheads, which are based on direct costs and lower capital charges, the annual revenue requirements of the double-alkali process are 12% lower than those of the limestone process.

ENERGY CONSUMPTION

A comparison of the total energy consumption for each process is shown in Table 20. Steam, diesel fuel, and electricity are shown in Btu equivalents.

The energy consumption in the sodium/limestone double-alkali process is only 71% of that for the forced-oxidation limestone process. Almost all of the difference is due to the significantly lower electrical consumption of the sodium/limestone double-alkali process.

TABLE 19. COMPARISON OF ANNUAL REVENUE REQUIREMENT COMPONENTS

	Annual revenue requirements, k$	
	Sodium/limestone double alkali	Limestone scrubbing
Raw materials		
Limestone	915	1,214
Soda ash	990	-
Lime	794	-
	2,699	1,214
Total labor	1,357	1,269
Utilities		
Steam	1,321	1,356
Electricity	1,096	2,133
Other	234	192
Maintenance	2,715	4,285
Landfill credit	(312)	-
Analysis	105	105
Direct costs	9,215	10,554
Overheads	2,319	3,395
Operating and maintenance costs	11,534	13,949
Levelized capital charges	13,987	15,151
Total first-year annual revenue requirements	25,521	29,100

TABLE 20. COMPARISON OF FGD PROCESS ENERGY CONSUMPTION

Process	Energy input, 10^9 Btu equivalent			% of plant capacity (gross)
	Electricity[a]	Steam[b]	Diesel fuel[c]	
Sodium/limestone double alkali	281.4	397.3	11.6[d]	2.6
Forced-oxidation limestone	547.7	407.7	14.9	3.7

a. 9,500 Btu/kWh.
b. 751.9 Btu/lb (based on recycle of condensate to power plant).
c. 144,000 Btu/gal.
d. Includes credit for fly ash disposal.

CONCLUSIONS AND RECOMMENDATIONS

The sodium/limestone double-alkali process is 8% lower in capital investment, 12% lower in first-year annual revenue requirements, and 14% lower in levelized annual revenue requirements than a comparable forced-oxidation limestone process. Although these economics may be only marginally attractive relative to limestone scrubbing (given the accuracy associated with this preliminary-grade economic evaluation), the sodium/limestone double-alkali process offers the potential for high SO_2 removal efficiencies at low absorber L/G ratios and for fewer maintenance problems with scaling, plugging, and erosion because calcium is primarily limited to the regeneration area.

The sodium/limestone double-alkali process is less energy intensive than a limestone-scrubbing process primarily because of the lower electrical consumption by recirculating pumps in the scrubbing area.

More development work on the sodium/limestone double-alkali process needs to be undertaken to demonstrate that both the design assumptions used in this study are achievable and the problems associated with a previous pilot-plant work at Scholz can be overcome with only minor (and relatively inexpensive) modifications. Both of these areas of uncertainty could be minimized with additional pilot-plant or prototype (3- to 30-MW equivalent) work which includes a long-term demonstration run. As a result of the unconfirmed design assumptions mentioned above, the estimated economics are considered potentially less accurate for the double-alkali process than for the forced-oxidation limestone process.

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